IN THE UNITED STATES PATENT AND TRADEMARK OFFICE

In Re Application of)	
)	FOR: LOW PRESSURE REVERSE
John E. TOMASCHKE)	OSMOSIS AND
)	NANOFILTRATION
Serial No.: 09/724,883)	MEMBRANES AND METHOD
)	FOR PRODUCTION THEREOF
Filed: November 28, 2000)	
)	Group Art Unit: 1723

DECLARATION OF JOHN TOMASCHKE UNDER 37 C.F.R. §1.132

Commissioner for Patents
Washington, D.C. 20231

      Attention: Examiner Krishnan Menon

I, JOHN E. TOMASCHKE, do hereby affirm and say that:

1.   I am an inventor of the above named application.

2.   The ESNA LF membranes discussed the references listed in the attached Appendices D, E and F are membranes made by Hydranautics, Inc., using the method of the invention.

3.   The first order of an ESNA LF membrane for testing was received by Hydranautics, Inc. on January 19, 2000. The element was shipped to the individuals performing the testing for the Boca Raton plant on January 28, 2000. To the best of my ability, this is the first case of public use that I have been able to identify.

DECYTOMA04.E05.frm

4.   The ESNA LF membranes were first offered for sale in a letter of intent to the

Deerfield Beach West Water Treatment Plant dated June 25, 2002.

5.   The ESNA LF membranes were offered for sale to the City of Boca Raton for use in

their water treatment facility that was the largest plant of its kind at the date of commission

late in 2003. The facility requires about 7690 membrane elements per year to purify about

40 million gallons per day. This is substantially more than most water treatment facilities at

that time that require about 192-960 membrane elements per year to purify about 1 to 5

million gallons per day.

6.   Multiple membranes provided by various manufacturers were tested for use in the

Boca Raton plant. One of the membranes tested was manufactured by Koch Membrane

Systems, Inc., the owner of the Chau US Patent No. 4,893,291. The chain of title is

attached as Appendix B. The membrane was found to be unsuitable for use in the Boca

Raton plant.

7.   No membranes were submitted by Saehan, the owner of the Koo US Patent

6,063,278. The chain of title is attached as Appendix G.

****

        I further declare that all statements made herein of my own knowledge are true and

that all statements made on information and belief are believed to be true; and further, that

these statements were made with the knowledge that willful false statements and the like

DECYTOMA04.E05.frm

so made are punishable by fine or imprisonment, or both, under Section 1001 of Title 18 of

the United States Code and that such willful false statements may jeopardize the validity of

the application or any patent issuing thereon.

Date: ___5/6/04___   Signed: ___John Tomarchl___

IN THE UNITED STATES PATENT AND TRADEMARK OFFICE

In Re Application of)	
)	FOR: LOW PRESSURE REVERSE
TOMASCHKE, John)	OSMOSIS AND NANOFILTRA-
)	TION MEMBRANES AND
Serial No.: 09/724,883)	METHOD FOR PRODUCTION
)	THEREOF
Filed: Novmeber 28, 2003)	
)	Group Art Unit: 1723

INFORMAL COMMUNICATION

Examiner Krishnan S. Menon
via Facsimile 571-273-1143
14 pages

Attention: Examiner Menon

Dear Examiner Menon:

Thank you for taking the time this morning to discuss the case with me and agreeing to the conference call on Friday. Below please find a proposed amendment and further remarks regarding the claims in the case. I look forward to speaking to you on Friday.

Please consider the following amendment:

IN THE CLAIMS

1-14. (cancelled)

15. (thrice amended) A composite membrane useful for reverse osmosis or nanofiltration comprising:

a supportive porous under-structure; and

a top layer consisting of a crosslinked polyamide thin film which is adhered to the upper surface of the porous support structure, said top layer having been contacted with a solution of a C_1-C_6 alkyl, alkenyl, haloalkyl, haloalkenyl, <u>or</u> hydroxy ~~or aryl~~ sulfonic or disulfonic acid compound,

whereby said membrane has a water flux of at least about 15 gfd and a rejection of at least 20 percent when tested an a 0.05 percent aqueous sodium chloride at 75 psi and 25°C.

16. (once amended) A composite membrane as in Claim 15 wherein said crosslinked polyamide comprises the reaction product of an aromatic diamine or triamine and an aromatic triacyl halide.

17. (once amended) A composite membrane as in Claim 16 wherein said crosslinked polyamide comprises the reaction product of an aromatic diamine or triamine, an aromatic triacyl halide and an aromatic diacyl halide.

18. (once amended) A composite membrane as in Claim 15 wherein the porous support is a polyarylethersulfone.

19. (once amended) A composite membrane as in Claim 15 wherein said first

membrane comprises a thin film, flat sheet, hollow fiber or tubular membrane.

20. (once amended) A composite membrane as in Claim 15 wherein the membrane is a component of a spiral-wound membrane filter or a plate and frame filter.

21. (thrice amended) A composite membrane as in Claim 15 wherein said organic sulfonic or disulfonic acid compound comprises a sulfoacetic, sulfosuccinic, ~~hydroxybenzene sulfonic, methane sulfuonic~~ methanesulfonic, ethanedisulfonic, or hydroxybutane sulfonic, ~~benzenedisulfonic, dihydroxy benzene sulfonic or dihydroxy benzene disulfonic~~ compound or mixtures thereof.

22. (cancelled)

23. (twice amended) A composite membrane as in Claim 15 wherein said organic sulfonic acid compound comprises methanesulfonic acid, trifluoromethanesulfonic acid or a mixture thereof.

24. (cancelled)

25. (once amended) A composite membrane as in Claim 15 wherein said solution of an organic sulfonic acid compound comprises said organic sulfonic acid compound dispersed or dissolved in water, alcohol, glycol, alkoxy alcohol or a carboxylic acid or a mixture thereof.

26. (cancelled)

[H:\Colleen\7703-hydranautics\PA02\informal communication04.A20.frm]

REMARKS

The Applicants thank the Examiner for the time and effort that he has put into this case in a number of telephonic interviews and looking for the file that was misplaced by the USPTO. The Applicants present a second informal communication in the case and a proposed amendment in order to try to bring the case to a conclusion with a notice of allowance.

In the Advisory Action mailed on July 11, 2003, the Examiner states that the reasons for the acid-treatment of Chau are not relevant in the context of a 102(b) rejection. All that matters is that Chau does an acid treatment. The Applicants are confused regarding the change of Chau from a 103 reference to a 102 reference. The Applicants request clarification if the Chau reference is now considered to anticipate the claims rather than making them obvious.

In a telephone conversation, the calculations done by the Examiner and subsequently by the Applicants in the previous informal communication were discussed. The Examiner stated that his calculations were rough calculations as opposed to the true numbers as calculated in the informal communication from the Applicants. The Examiners calculations provided a number of about 13.2 gfd with a pressure of 75 psig. The Examiner stated that when this number was further adjusted to account for the difference in salt concentration of 500 ppm, as recited in claim 15, as compared to 2000 ppm as in Chau, the flux would be greater than 15 gfd as required by the claims. However, no calculations are presented. Therefore, it is difficult to determine if the calculation was made correctly. The Applicants have calculated that the flux of the PTSA treated membrane of Chau would be about 13.57 after having adjusted for the difference in salt concentration, not before. The Applicants request that the Examiner provide his calculations to the Applicants to determine who is in error.

[H:\Colleen\7703-hydranautics\PA02\informal communication04.A20.frm]

In a telephone conversation, the Examiner stated that he believed that about 13 gfd was within the range of the breadth of claim 15 that recites "about at least 15 gfd." The Applicants submit that those skilled in the art would not consider about 13 gfd to be "about at least 15 gfd" as stated in the claims. The Applicants request that the Examiner demonstrate that variation of about 10% is considered to be within the variation of "about" in the context of the instant Application. It is noted by the Applicants that both the Chau patent and the instant application provide data to a tenth gfd.

The Applicants have proposed an amendment to claim 15 as shown above. The amendment eliminates the use of aryl sulfonic acids in the method of the invention. Claim 21 has also been amended to eliminate references to aryl sulfonic acids. In view of this amendment, and the previous informal communication, the Chau reference can no longer be considered to teach each element of the claims. A rejection under 35 U.S.C. 102 requires that "each and every element as set forth in the claim is found, either expressly or inherently described, in a single prior art reference." *Verdegaal Bros. v. Union Oil Co. of California,* 814 F.2d 628, 631, 2 USPQ2d 1051, 1053 (Fed. Cir. 1987). Based on the calculations and remarks presented in the previous informal communication, the only membranes of Chau that have sufficiently high flux to meet the requirements of claim 15, are made by treating the membrane with 2-10% citric acid. Citric acid is not a sulfonic or disulfonic acid as required by claim 15. Therefore, claim 15 cannot be rejected for anticipation under 35 U.S.C. 102(b) in view of Chau, if such a rejection was intended by the Examiner. The rejection is traversed.

If a rejection of the claims can be made, the rejection must then be under 103(a) as stated by the Examiner in the office action of March 26, 2003. In the Advisory Action, the Examiner states that the Chau reference teaches all of the

[H:\Colleen\7703-hydranautics\PA02\informal communication04.A20.frm]

elements of the instant invention and the Koo reference simply provides a list of sulfonic acids that one might use in RO membrane formation. The Applicants submit that the Chau reference teaches a large number of compounds that can be used for the post-treatment of RO membranes to allow them to be dried between manufacture and use. The list of sulfonic acids provided by Koo are for membrane preparation, not post-treatment. The steps are not equivalent; therefore, compounds that can be used for one step cannot necessarily be used for a different step in another application. Therefore, the combination of references is improper as previously discussed.

The Chau reference teaches a variety of compounds for post-treatment of membranes. The list given in the abstract includes acid and amine salts selected from the group consisting of hydroxy polycarboxylic acid, polyaminoalkylene polycarboxylic acid, sulfonic acid, amine salts of acids, amino acid, amino acid salt, polymeric acid and inorganic acid. The compounds have many dissimilar characteristics, with a common characteristic of all being hygroscopic. No preference for any particular compound is taught by the Chau reference. Therefore, all of the compounds are considered to be equivalent.

In the MPEP, section 2144.08(II) it states that rejection of a claims under 103(a) "to a specific compound or subgenus embraced by a prior art genus should be analyzed no differently than any other claims for the purposes of 35 U.S.C. 103.... The fact that a claimed species or subgenus is encompassed by a prior art genus is not sufficient by itself to establish a *prima facie* case of obviousness.. *In re Baird*, 16 F.3d 380, 382, 29 USPQ2d 1550, 1552 (Fed. Cir. 1994)("The fact that a claimed compound may be encompassed by a disclosed generic formula does not by itself render that compound obvious.") *In re Jones*, 958 F.2d 347, 350, 21 USPQ2d 1941, 1943 (Fed. Cir. 1992)"

Further, in determining whether the claimed species or subgenus would have been obvious to one of ordinary skill in the art at the time of the invention, if the motivation to select the claimed species or subgenus must be provided by the reference (MPEP2144.08(II)(A)(4)). The considerations regarding the motivation to select a specific species includes the size of the genus, the express teachings of the reference, the teachings of structural similarity, the teachings of similar properties, the predictability of the technology and any other teachings to support the selection of the species or subgenus. The Applicants will consider these points individually.

The size of the genus of compounds for use in the Chau invention is substantial. The following classes are listed hydroxy polycarboxylic acid, polyaminoalkylene polycarboxylic acid, sulfonic acid, amine salts of acids, amino acid, amino acid salt, polymeric acid and inorganic acid. The Applicants have included two excerpts from the Columbia Encyclopedia to demonstrate the breadth of two of the classes listed, inorganic acids and sulfonic acids. Four common inorganic acids are listed in the excerpt on acids and bases, suggesting far more less common inorganic acids. A number of classes of sulfonic acids are listed in the excerpt on sulfonic acids. Each one of the eight classes taught by Chau would be too large to search through for compounds that were functional in the instant invention as claimed. Moreover, in Chau there is no suggestion of how or where to break the large classes into smaller groups. There is no suggestion of a difference between any of the compounds of the classes listed. There is no suggestion that there may be subgroups within the classes that may have different properties. In the instant invention, it was determined that lower molecular weight sulfonic acids (C_1-C_6) as claimed were preferred for use in the instant invention. This class is not offered as an option in the Chau reference; therefore, there can be no motivation to

elect the group based on the Chau reference.

There are no express teachings in the Chau reference that the class of C_1-C_6 sulfonic acids even exists. There are only teachings directed to the large class of sulfonic acids. Therefore, no motivation exists to select the specific compounds claimed in the instant invention.

The genus of compounds taught by Chau are not structurally similar to each other. No preference is taught for the use of any one structural class of compounds over the others. Within the classes taught by Chau, there are no distinctions made based on molecular weight or number of carbons. No compounds that fall within the class claimed in the instant invention were tested by Chau. The p-touluenesulfonic acid taught by Chau is a C_7 sulfonic acid. No teachings regarding a structurally desirable compound for the treatment of membranes is provided by Chau. Therefore, no motivation exists to select the compounds claimed in the instant invention based on structure.

The genus of compounds taught by Chau have substantially different properties, with the one common feature of all being hygroscopic. There is no grouping of compounds by properties in Chau. Therefore, one cannot be motivated to select a group based on properties.

The substantial amount of data in both the instant application and the patent of Chau suggest a substantial level of non-predictability in the art. Therefore, one could not predict that the C_1-C_6 sulfonic acids claimed in the instant invention would result in the high flux membranes of the instant invention.

The Applicants could find no other teachings in the Chau reference that could be considered to suggest the claimed group as taught by the instant invention.

In regard to the teachings of the Koo reference of sulfonic acids that can be used in the preparation of RO membranes, a long series of sulfonic acids including

[H:\Colleen\7703-hydranautics\PAO2\informal communication04.A20.frm]

methanesulfonic acid, toluenesulfonic acid, camphorsulfonic acid, ethanesulfonic acid and benzenesulfonic acid, or any other aromatic, aliphatic, or cycloaliphatic sulfonic acid are listed. The majority of these sulfonic acids do not fall within the limitations of the claims as they have more than 6 carbons. Toluene is C_7. Camphor is C_{10}. Benzene sulfonic acid is an aryl compound. Aliphatic compounds may be more than 6 carbons, and it is likely that cycloaliphatics are more than six carbons to allow ring formation. Therefore, although Koo does teach a large number of sulfonic acids that can be used in RO membrane preparation, very few can be used with the instant invention.

Moreover, the sulfonic acids are grouped with any of a number of strong acids that serve as a reactant to make a salt with a polyfunctional amine (col 3, ln 3-5). The acid itself is not applied to the membrane, instead the membrane is exposed to the reaction product of the strong acid. Being a strong acid in an aqueous solution, the acid is dissociated and no longer a compound that could be used in the instant application. The selection of the specific subgenus of C_1-C_6 sulfonic acids as claimed by the instant invention from the variety of sulfonic acids taught by Koo cannot be made based on any of the teachings of Koo. Therefore, the reference is again lacking in the specific teachings that would be required to make the instant invention obvious.

The broad teachings of the prior art do not distinguish a group of C_1-C_6 sulfonic acids. They only teach sulfonic acids as a large, homogeneous class as in Chau, or as a large group of sulfonic acids with no distinctions based on size as in Koo. Claim 15 of the instant invention, the only independent claim pending in the instant application, teaches a specific subclass of sulfonic acids. The existence of this subclass is neither taught nor suggested by the prior art. Therefore, the unique

properties of this subclass cannot be taught or suggested by the prior art. As all of the remaining claims pending in the case are dependent either directly or indirectly on claim 15, they are also clearly distinguished from the prior art. Therefore, the rejection of all of the claims for obviousness under 35 U.S.C. are traversed.

CONCLUSIONS

The Applicants thank the Examiner for his time and consideration and look forward to a formal telephonic interview.

Respectfully submitted,

Dated: January 22, 2004   By: _____

Colleen J. McKiernan
Agent for Applicant
Registration No. 48,570

GORDON & REES, LLP
1660 Union Street
San Diego, California 92101

Telephone: (619) 238-0999
Direct Line: (619) 702-1967
Facsimile: (619) 238-0062

Docket No.: 7703-PA02

IN THE UNITED STATES PATENT AND TRADEMARK OFFICE

In Re Application of	)
	) FOR: LOW PRESSURE REVERSE
TOMASCHKE, John	) OSMOSIS AND NANOFILTRA-
	) TION MEMBRANES AND
Serial No.: 09/724,883	) METHOD FOR PRODUCTION
	) THEREOF
Filed: Novmeber 28, 2003	)
	) Group Art Unit: 1723

INFORMAL COMMUNICATION

Examiner Krishnan S. Menon
via Facsimile 703-746-8788

Attention: Examiner Menon

Dear Examiner Menon:

As we discussed by telephone, below please find some calculations and issues relevant to the prosecution of this case. When you have had an opportunity to discuss this information with your supervisor, please contact me by telephone or e-mail so that we may discuss the case. I hope that we will be able to resolve some of the outstanding issues in the case so that the prosecution may close with the issuance of a notice of allowance.

Bridging pages 3-4 of the office action mailed on July 11, 2003, you that upon normalizing the data of Chau to allow for comparison to the conditions in claim 15, the membrane of Chau would meet the requirements of the claim. The Applicants submit that an error was made in the calculation by the Examiner and that essentially none of the membranes taught by Chau would have sufficient flux

or salt rejection as required by claim 15. The Applicants concur that the reduction of the salt concentration in the instant invention reduces osmotic pressure. However, the instant invention also requires a substantially lower operating pressure, 75 psi as compared to 220 psi which substantially reduces the net driving pressure, decreasing flux.

Below please find calculations performed to normalize the data to the limitations of claim 15 in the instant application. An explanation of the calculations made is presented.

The following equations are used in the comparison of flux at different salt concentrations and operating pressures.

Net Driving Pressure (NDP) = Operating Pressure - Osmotic Pressure

Salt passage = 100 - % rejection

Osmotic pressure = salt concentration in ppm x 0.0115

In the PTSA treated membrane of Chau in Example 5, Table 4:

Osmotic pressure = 2000 x 0.0115
= 23 psi

Operating pressure = 220 psi

Net Driving Pressure = 220 psi -23 psi
= 197 psi

In Claim 15 Osmotic pressure = 500 x 0.0115
= 5.75 psi

Operating pressure = 75 psi

Net Driving Pressure = 75 psi - 5.75 psi
= 69.25 psi

[H:\Colleen\7703-hydranautics\PA02\informal communication.frm]

The NDP factor $= NDP_{Chau}/ NDP_{Claim\ 1}$
                  $= 197/69.25$
                  $= 2.845\ x$

Therefore, the flux of the PTSA treated membrane of Chau under the conditions of claim 15 is:

$$38.6/2.845 = 13.57\ gfd$$

This is less than the 15 gfd as required in claim 1.

The salt passage would also be altered by the same NPD factor. Therefore, under the conditions of claim 15, the salt passage of Chau would be:

$$1.8 \times 2.845 = 5.12;\qquad 100 - 5.12 = 94.88\%\ salt\ rejection.$$

All of the testing of Chau was performed under a single set of conditions. The Applicants have calculated the flux for each of the membrane conditions and found that only three of the membranes of Chau have sufficiently high flux to meet the requirements of claim 1 in regard to both flux and salt rejection. They are membranes D-2, D-3 and D-4 in Table 2. The membranes have fluxes between 16.0 gfd and 23.44 gfd and salt rejections between 94.0% and 83.5%. Although these membranes meet the working specifications of Claim 1, they are prepared by treating the membrane with 2-10% citric acid. It is of note that the membranes prepared by the method of claim 15 have higher salt rejections at comparable fluxes. Therefore, the rejection of claim 1 under 35 U.S.C. 102(b) is traversed.

The Applicants further state that in view of the large number of compounds tested by Chau that did not produce membranes that meet the specifications of claim 15, it would not have been obvious to one skilled in the art to test the specific class of compounds claimed, C_1-C_6 sulfonic acids. Therefore, claim 15 is
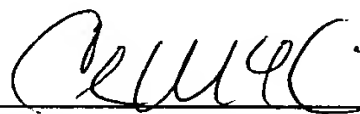
not obvious in view of Chau and a rejection under 35 U.S.C. 103 is not appropriate.

CONCLUSIONS

The Applicants thank the Examiner for his time and consideration and look forward to a formal telephonic interview.

Respectfully submitted,

Dated: _Oct 15, 2003_

By: _____
Colleen J. McKiernan
Attorney for Applicant
Registration No. 48,570

BROWN MARTIN HALLER & McCLAIN LLP
1660 Union Street
San Diego, California 92101

Telephone: (619) 238-0999
Facsimile: (619) 238-0062
E-mail: colleen@bmhm.com

Docket No.: 7703-PA02

United States Patent and Trademark Office

Home | Site Index | Search | Guides | Contacts | eBusiness | eBiz alerts | News | Help

Assignments on the Web > **Patent Query**

Patent Assignment Abstract of Title
NOTE:Results display only for issued patents and published applications. For pending or abandoned applications please consult USPTO staff.

Total Assignments: 3
Patent #: 4983291   **Issue Dt:** 01/08/1991   **Application #:** 07450637   **Filing Dt:** 12/14/1989
Inventors: MICHAEL M. CHAU, WILLIAM G. LIGHT, HONG C. CHU
     Title: DRY HIGH FLUX SEMIPERMEABLE MEMBRANES
Assignment: 1
     Reel/Frame: 005281/0231   **Recorded:** 04/25/1990   **Pages:** 3
     Conveyance: ASSIGNMENT OF ASSIGNORS INTEREST.
          Assignors: CHAU, MICHAEL M.   **Exec Dt:** 12/12/1989
                    LIGHT, WILLIAM G.   **Exec Dt:** 12/12/1989
                    CHU, HONG C.   **Exec Dt:** 12/12/1989
          Assignee: ALLIED-SIGNAL INC., COLUMBIA ROAD AND PARK AVE., MORRIS TOWNSHIP, NJ A CORP. OF DE
Correspondent: HAROLD N. WELLS
                    ENGINEERED MATERIALS RESEARCH CENTER
                    50 EAST ALGONQUIN ROAD
                    DES PLAINS, IL 60017-5016
Assignment: 2
     Reel/Frame: 006950/0861   **Recorded:** 04/22/1994   **Pages:** 2
     Conveyance: ASSIGNMENT OF ASSIGNORS INTEREST (SEE DOCUMENT FOR DETAILS).
          Assignor: ALLIEDSIGNAL INC., (FORMERLY ALLIED CORPORATION)   **Exec Dt:** 03/31/1994
          Assignee: FLUID SYSTEMS CORPORATION
                    10054 OLD GROVE ROAD, SAN DIEGO, CA
Correspondent: TERI L. GLASER
                    LEGAL ASSISTANT
                    WHITE & CASE
                    1155 AVENUE OF THE AMERICAS
                    NEW YORK, NY 10036
Assignment: 3
     Reel/Frame: 009781/0880   **Recorded:** 03/02/1999   **Pages:** 2
     Conveyance: CHANGE OF NAME (SEE DOCUMENT FOR DETAILS).
          Assignor: FLUID SYSTEMS CORPORATION   **Exec Dt:** 01/31/1999
          Assignee: KOCH MEMBRANE SYSTEMS, INC.
                    4111 EAST 37TH STREET NORTH
                    WICHITA, KANSAS 67220
Correspondent: SHOOK, HARDY & BACON L.L.P.

JOAN OPTICAN HERMAN
1200 MAIN STREET
KANSAS CITY, MO 64105-2118

Search Results as of: 05/07/2004 12:55 PM

If you have any comments or questions concerning the data displayed, contact OPR / Assignments at 703-308-9723

| .HOME | INDEX| SEARCH | eBUSINESS | CONTACT US | PRIVACY STATEMENT

Optimizing the Performance of Low Fouling Membranes for the World's Largest Nanofiltration Plant

by

Author: Curtis A. Kiefer – Membrane Treatment Specialist
Camp, Dresser & McKee, Fort Lauderdale, FL

Co-Author: Frank Brinson – Project Manager
Camp, Dresser & McKee, Fort Lauderdale, FL

Co-Author: Bill Suratt – Consulting Division
Camp, Dresser & McKee, Fort Lauderdale, FL

Abstract

The City of Boca Raton Florida is currently constructing a 40-mgd nanofiltration plant. This will be the largest plant of its kind in the world when it is commissioned in late 2003. This paper focuses on the design features of the plant and the results of extensive pilot and bench-scale tests. The pilot and bench-scale tests were conducted to find ways to reduce the severe fouling caused by the very high organic content of the raw ground water. Some interesting results from the testing are discussed including zeta potential and electrophoretic testing and side-by-side comparisons of various membranes and antiscalants.

Introduction

In order to reduce finished water color levels and enhance their ability to meet drinking water regulations for disinfection by-products (DBP), the City of Boca Raton is in the process of constructing a 40-mgd nanofiltration plant—the largest of its kind in the world.

In recent years, the City of Boca Raton, Florida, has experienced a gradual, but steady increase in the levels of color in the raw and finished water at its Glades Road Water Treatment Plant. In addition, high levels of dissolved organics in the raw water have made compliance with new, more stringent regulations for DBPs more difficult with the existing lime softening process. While the conventional lime process does remove some color, DBP precursors, and associated constituents, it is not capable of consistently and reliably meeting both DBP and color standards simultaneously. Faced with increasing customer dissatisfaction with high color levels in the finished water and stricter drinking water regulations for DBPs, the City recognized the need to pursue other treatment process alternatives to meet its drinking water quality objectives.

In order to meet these goals, the City evaluated several alternatives for process modifications, including enhanced lime softening, ozonation, and membrane softening. Due to the efficiency of the membrane process in simultaneously removing color and DBP precursors, the City selected the membrane softening process to enhance their existing treatment facilities. This technology is being used increasingly in Florida as an

alternative (or a complement) to conventional lime softening to treat relatively high colored groundwater supplies such as the City's. Nanofiltration membranes provide a way of removing over 90 percent of the total organic carbon (TOC) from the raw water therefore eliminating the majority of the precursors to DBPs. In addition, pilot testing of nanofiltration at the City produced finished water with a color of less than two color units (CU).

The Camp Dresser & McKee (CDM) project team includes CH2M Hill and laboratory research on membrane fouling has been conducted by Dr. Harvey Winters.

Early Pilot Testing

Early in the design phase of this project, it was recognized that the raw water at the Glades Road Water Treatment Plant (WTP) presented several challenges for developing a successful membrane treatment process. A membrane pilot test program was initiated to address pretreatment concerns and to develop a membrane treatment process design that would meet the City's water treatment objectives. The pilot testing revealed that cartridge filter and membrane fouling was a significant problem that needed to be addressed in the design. Fouling issues included sand and silt, oxidation of iron and hydrogen sulfide, biofouling, and dissolved organic foulants. Due to the extent of fouling encountered in the pilot testing, it was evident that the raw water quality must be improved through rehabilitation of the existing wellfield and/or installation of multimedia pressure filters. Initial fouling of the cartridge filters was so rapid that filter cartridge replacement was required approximately every seven days. To address this problem, a multimedia filter was installed upstream of the cartridge filters to lengthen their run time.

After the installation of the multimedia filter, a significant improvement was seen in the operating life of the cartridge filters; however, membrane fouling continued to be an issue. The pilot plant operating data still showed a continuous decline in membrane flux due to dissolved organic foulants. These flux decline trends in all three stages of the membrane pilot plant are shown in **Figure 1**.

A variety of pretreatment chemical combinations and membrane models were investigated in an effort to alleviate these fouling problems. Pilot testing activities aimed at addressing this membrane fouling problem included:

- Demonstration of operation without acid addition
- Side-by-side comparisons of several membranes with respect to fouling characteristics and hardness rejection
- Side-by-side comparisons of antiscalants and dispersants to assess impacts on fouling

The results of these tests are summarized in **Figure 1**. During the operating period from 25500 to 30800 hours, four test runs using different combinations of acid and antiscalants were conducted. These tests included several brands of the most commonly used antiscalants and antifoulants in the industry. These tests were characterized by membrane fouling and rapid declines in membrane mass transfer coefficient. Fouling at

these rates would require cleaning at less than three month intervals, which was considered to be unacceptable.

## FIGURE 1. MASS TRANSFER COEFFICIENT VS. TIME OF OPERATION
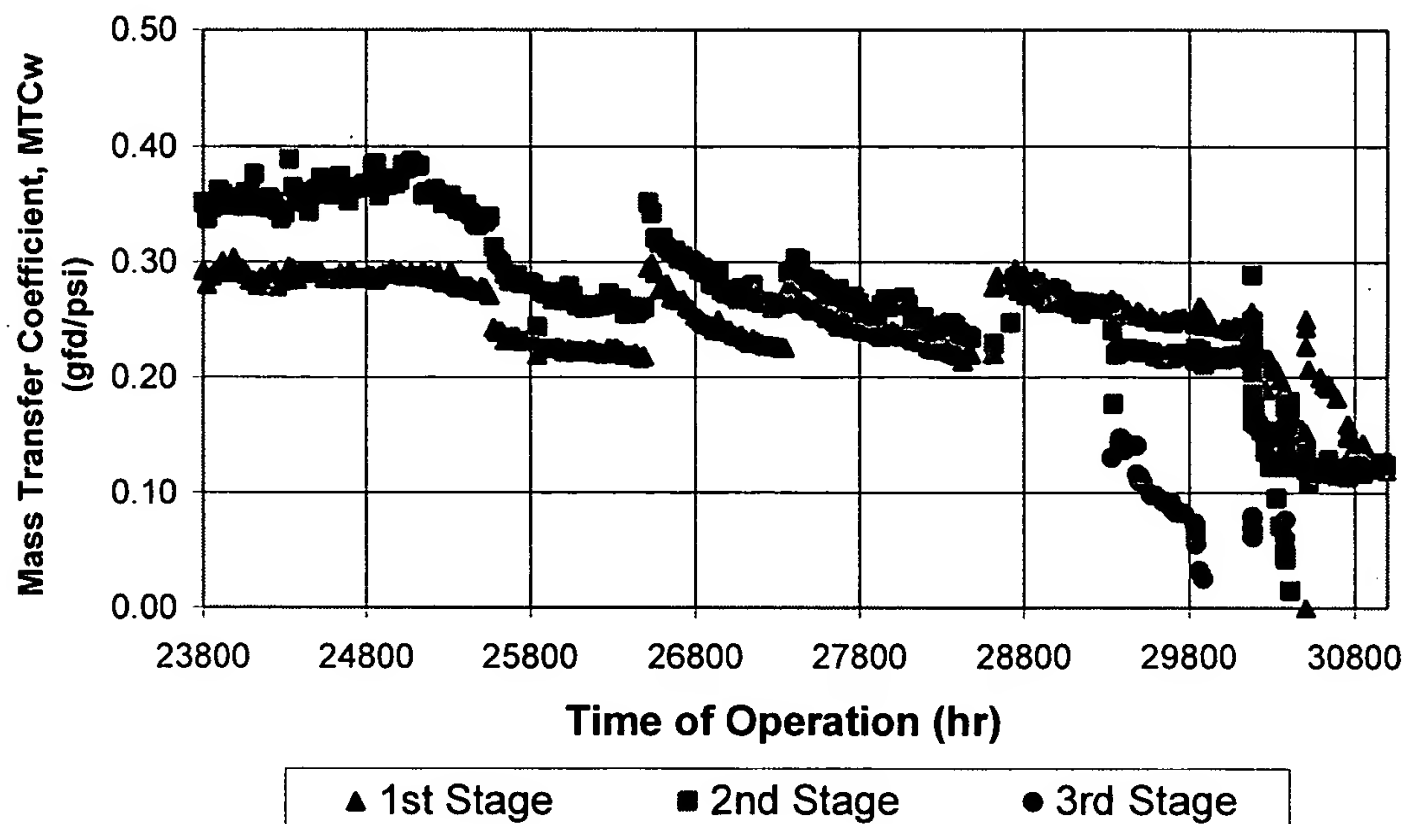
Operation without Acid and Antiscalant

After several unsuccessful test runs using some of the most widely used pretreatment chemicals in the membrane industry, several membranes were removed from the pilot test unit for autopsy analysis. The results of this testing revealed that, while substantial fractions of the foulants were consistent with naturally occurring humic acids, other components consistent with the active ingredients of the antiscalant chemicals were also observed in significant concentrations. The results of the membrane autopsies and observations from other membrane treatment plants in South Florida with high humic acid concentrations appeared to indicate that these antiscalants may actually be complexing with the naturally occurring humic acids and may be contributing to the fouling problem. To test this theory, it was decided to operate the pilot unit without either acid or antiscalant addition.

A comparison of operation with and without antiscalant addition is illustrated in **Figure 2.** The initial 20 days of operation is indicative of pilot unit performance using antiscalant with no acid addition. Operation during this period is characterized by fouling and a rapid decline in mass transfer coefficient in both stages. Testing from day 22 through the remainder of the test run was conducted without acid or antiscalant pretreatment chemicals. Although there was slightly more than a 10 percent decline in

mass transfer coefficient in the second stage, operation during this period was much more stable than in the previous test and the testing summarized in **Figure 1**.

## FIGURE 2. MASS TRANSFER COEFFICIENT VS. TIME OF OPERATION
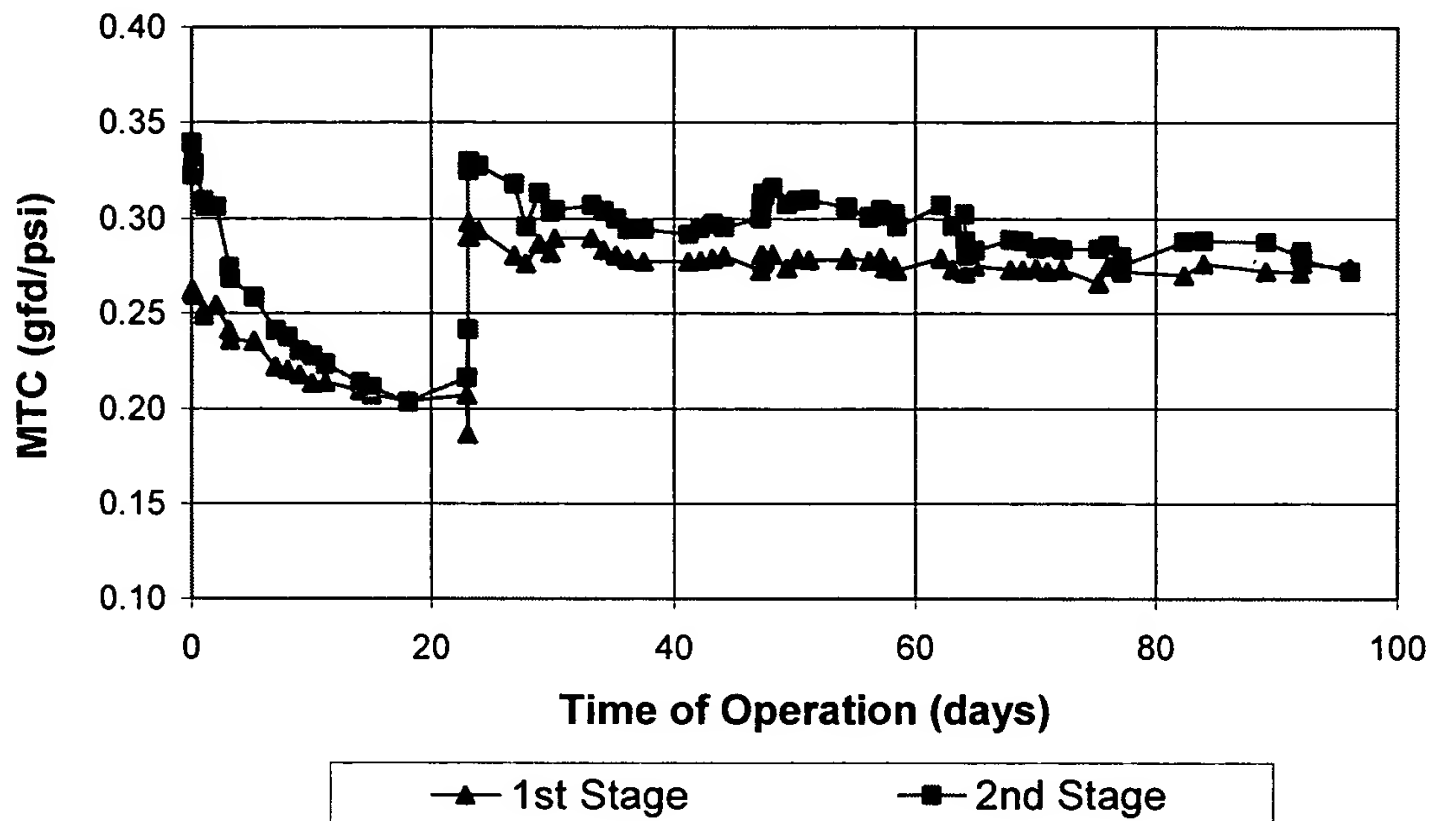
New Low Fouling Membranes

Following the initial successful trial run without acid and antiscalant, it was decided to replace the membranes, which had been installed in the pilot unit during the various pretreatment chemical trial runs, with a new set of membranes. The membranes selected for this testing were a new prototype model from Hydranautics that were developed specifically to have a reduced fouling potential for waters with high humic acid concentrations. These membrane elements were predecessors to what would later become the ESNA LF line of low fouling nanofiltration membranes. The ESNA LF membranes will use similar low fouling technology, which has already been applied to their LF series of membrane for low pressure RO applications.

The results of the first sixty days of testing with the new low fouling membranes are presented in **Figure 3**. A minor adjustment was made to the pilot unit operation between day 17 and 20; however, it can be seen that the mass transfer coefficients during this period of operation were very stable. These results represented the best performance with respect to fouling that had been achieved with the pilot testing up to that point.

The pilot test unit continued to operate with the ESNA LF membranes for approximately six to eight months without acid and antiscalant addition. Operation during this period was generally stable with respect to humic acid fouling; however, there were periods in which SDI's from the wellfield were higher than recommended for pilot unit operation.

4

FIGURE 3. ESNA LF MASS TRANSFER COEFFICIENT VS. TIME

City of Boca Raton Membrane Pilot Plant
Hydranautics ESNA LF
Mass Transfer Coefficient
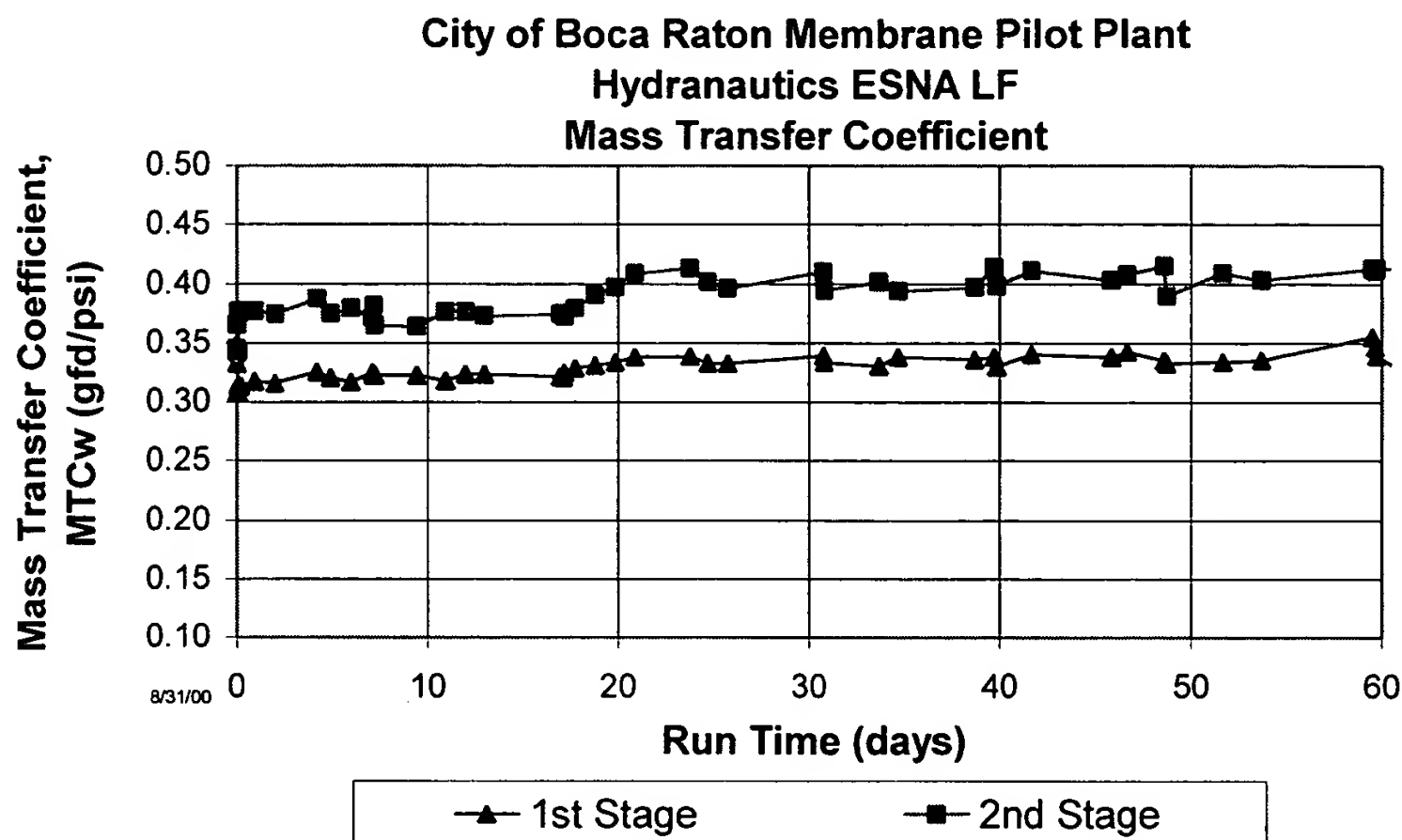
This problem was compounded by a drought situation that resulted in water use restrictions. The drought lowered groundwater levels, which resulted in entrainment of air in some wells with marginal submergence over the pump impellers. In addition, the water use restrictions resulted in frequent cycling of the wellfield between watering and non-watering days. Both the lower groundwater levels and more frequent cycling of the wellfield resulted in increased air entrainment and possibly more solids production from the wells. Due to the presence of hydrogen sulfide and iron in the raw water, this air entrainment resulted in increased formation of colloidal sulfur and ferric hydroxide and unacceptable SDI readings.

The problem with high SDI's was resolved by providing a new raw water connection for the membrane pilot unit. This new raw water connection allowed the pilot unit to operate exclusively on water from the northwest wellfield as opposed to a composite blend of all wellfields. The northwest wellfield will now be the primary raw water supply for the new full-scale membrane plant; therefore, this configuration is considered to be more representative of actual full-scale plant operating conditions. After the installation of the new raw water line, SDI's were reduced to the range of 2.0 to 2.5 and allowed pilot testing to resume.

Side-by-Side Testing

Having demonstrated satisfactory operating performance without acid and antiscalant and having resolved the high SDI problem, it was decided that membranes from three major

5

membrane manufacturers should be tested side-by-side in the pilot test unit. The primary objectives of this testing were:

- Verify that the project water quality objectives could be achieved and specifically determine that the membranes would operate within the required applied pressure criteria and would produce water in the proper hardness range.

- Confirm that membranes from approved membrane manufacturers could perform satisfactorily without acid and antiscalant addition.

- Confirm that membranes from approved membrane manufacturers perform in accordance with their performance projection programs.

The primary permeate water quality requirements for the membrane treatment plant design with the corresponding raw water values are summarized in Table 1. As discussed

TABLE 1 – REQUIRED WATER QUALITY

Parameter	Unit	Projected Raw Water	Permeate Quality Stage 1&2
Bicarbonate*	Mg/L	265	<175
Color	CU	50	<2.0
Sum of Ions	Mg/L	450-500	<300
Total Hardness	Mg/L as $CaCO_3$	250	50-80
Total Organic Carbon	Mg/L as C	12.0	<1.0
Total Trihalomethane** Formation Potential	Mg/L	0.60	<0.042
Haloacetic Acid Formation Potential***	Mg/L	0.40	<0.030

in the introduction, primary water quality parameters of concern for this project include color, hardness and disinfection byproduct formation potential. The blending rate of lime softened water with membrane softened water was selected to result in a finished water color of approximately 5.0. Since the lime softened water typically averages 12 CU, a 2:1 blending rate of membrane softened to lime softened water will result in the required finished water color. For corrosion control purposes the City prefers to maintain finished water in a hardness range between 70 and 90 mg/l as $CaCO_3$. Based on the selected blending ratio and typical lime softened water hardness, the permeate from the membrane softening facility should have a hardness in the range of 50 to 80 mg/l as $CaCO_3$ to meet this objective. Similarly permeate THM and HAA formation potential need to be within the specified ranges to satisfy the finished water quality requirements.

Three major nanofiltration membrane element manufacturers were asked to provide membrane performance projections for the full-scale membrane plant and seven of the corresponding model of 4-inch diameter membrane elements for the side-by-side testing. Based on a review of these membrane performance projections, the maximum design transmembrane pressure for this project was set at 80 psi. In general, elements from all three manufacturers were able to operate within the required transmembrane pressure.

FIGURE 4. ESNA LF PERMEATE HARDNESS VS. TIME

**City of Boca Raton Pilot Plant**
Hardness
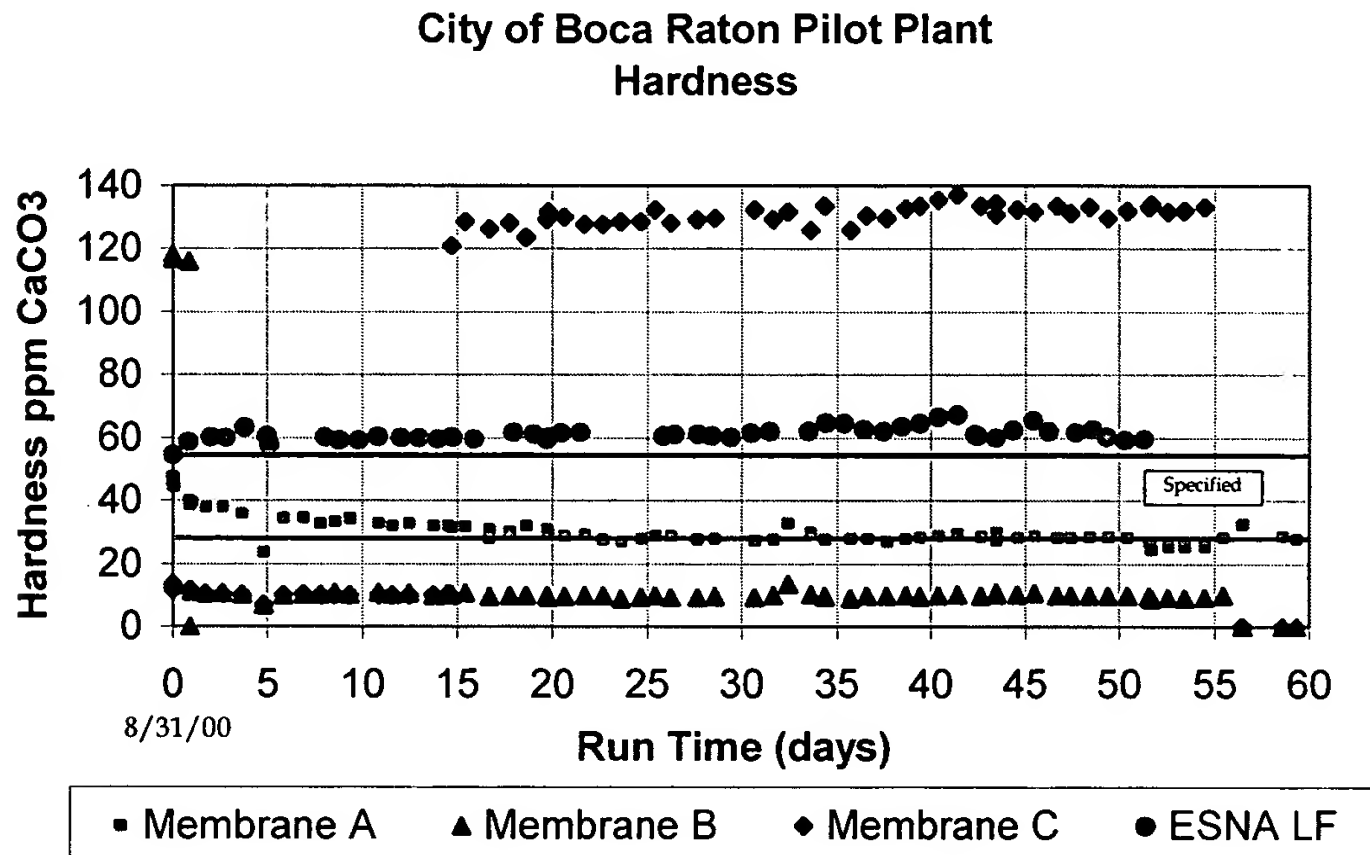
In contrast, only one of the membrane element manufacturers was able to produce permeate in the specified hardness range. A graph of permeate hardness versus time for the side-by-side testing is provided in **Figure 4**. This graph illustrates that the membranes initially provided by both membrane manufacturers B and C produced permeate with hardness well below the specified range. Manufacture C subsequently provided an alternate membrane element with lower hardness rejection. These new membrane elements were installed in the pilot unit on day 15 and as shown in the graph, the permeate hardness produced by this membrane was well above the desirable range. Neither manufactures B or C were able to offer hybrid designs acceptable to the City which could meet the permeate hardness target.

Only the membrane from manufacturer A produced water in the specified hardness range. Although the ESNA LF membrane was not originally included in the side-by-side testing due to price considerations, this membrane was subsequently added to the testing because of its more stable operating performance. As shown in **Figure 4** although membrane A produced permeate hardness in the proper range, the membrane exhibited more susceptibility to humic acid fouling as evidenced by decreasing hardness passage with time.

The fouling tendency of membrane A is further illustrated in the plot of mass transfer coefficient vs time presented in **Figure 5**. In this side-by-side test, the mass transfer coefficient of membrane A decreased by approximately 33 percent. In contrast, the mass transfer coefficient for ESNA LF membrane was very stable and actually exhibited a slight increase over the first few weeks of operation.

FIGURE 5. MASS TRANSFER COEFFICIENT VS. TIME

City of Boca Raton Pilot Plant
**Mass Transfer Coefficient**
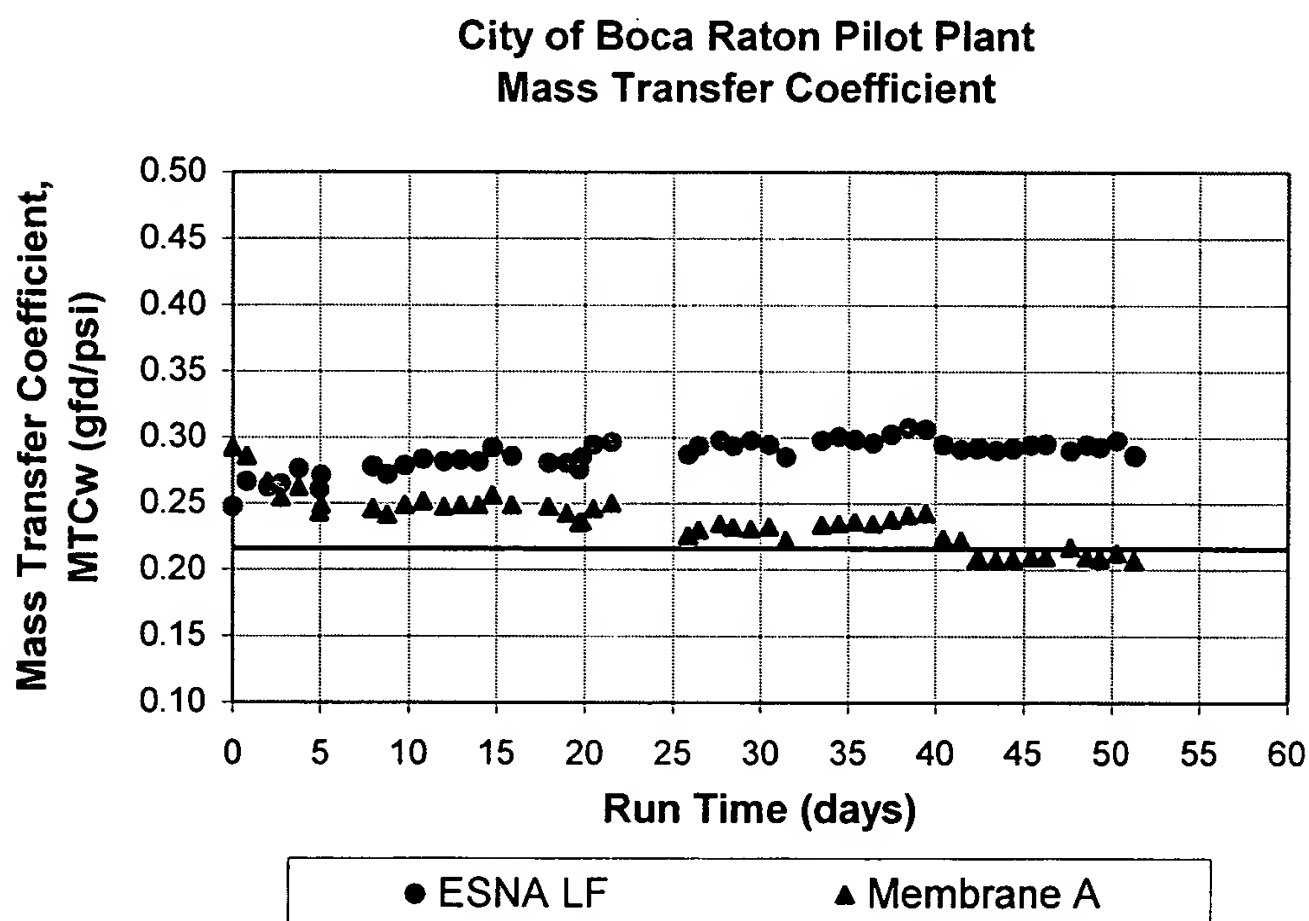
2:1 Array Testing

As shown in **Figure 5** the permeate hardness of the ESNA LF membranes which were originally tested, produced permeate with a hardness just above the desirable hardness range. Through the course of the side-by-side testing, Hydranautics exhibited greater flexibility in optimizing the performance of the ESNA LF to meet the permeate quality requirements. A series of ESNA LF membranes were produced with hardness rejections of 92, 95, and 98 percent. To demonstrate that the ESNA LF membrane could be made to meet the required permeate hardness range for this project, a 2:1 membrane array was tested with 95 percent hardness rejection ESNA LF membranes in the first stage and 92 percent hardness rejection membranes in the second stage.

The results of the 2:1 array testing are shown in **Figure 6.** The graph illustrates the permeate hardness for the first stage, second stage and composite total permeate hardness. The composite permeate hardness for the 2:1 array was approximately 60 mg/l which is close to the middle of the desirable permeate hardness range.

FIGURE 6. PERMEATE HARDNESS VS. TIME

City of Boca Raton Pilot Plant
2:1 Array Hardness

▲ Stage 1   ▪ Stage 2   ◆ COMPOSITE
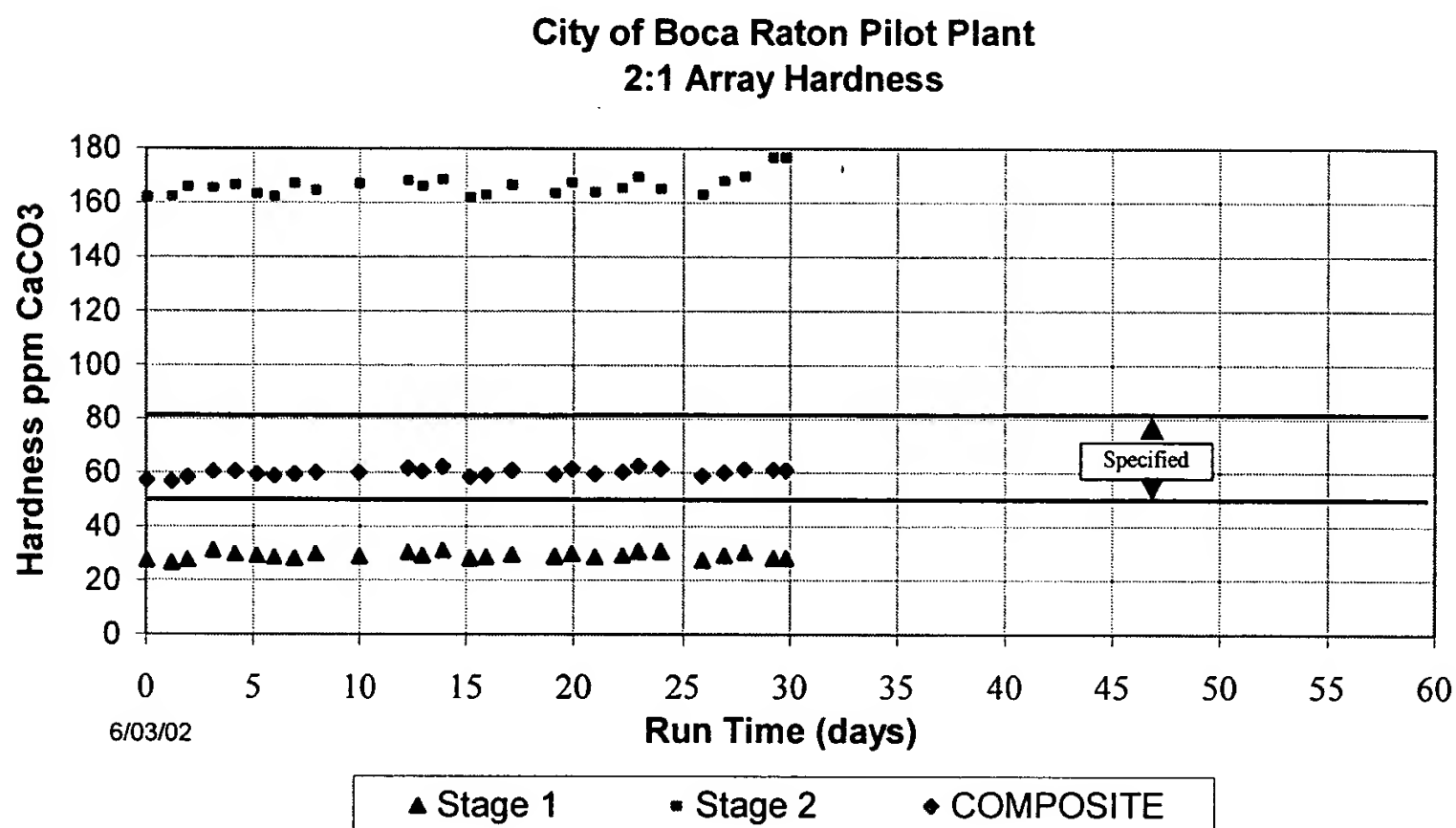
Summary

Although numerous problems were encountered in developing the design for the largest nanofiltration plant in the world, through a comprehensive pilot testing program these problems were overcome and an optimum membrane treatment solution was developed. Initial severe colloidal fouling problems were solved through a combination of wellfield rehabilitation and installation of multimedia filtration. When traditional chemical pretreatment options failed due to unacceptably high fouling rates associated with naturally occurring organics, an unconventional approached was adopted that eliminated pretreatment chemicals and put the naturally occurring organics to work as scale controlling agents. Not only does this approach simplify operation and avoid potential problems associated with acid storage and handling, but more importantly it significantly reduces operating costs. Initial calculations indicated that this treatment facility would require one tanker truckload of acid per day at a cost of approximately $3,000 per day. In addition to cost savings associated with eliminating acid, similar savings will also be realized due to the elimination of antiscalant.

Another revolutionary development, which evolved through this pilot test program, was the introduction of low fouling technology to nanofiltration membranes. The viability of

9

long-term operation without acid and antiscalant was first demonstrated using the new low fouling ESNA LF nanofiltration membranes. These membranes demonstrate much less susceptibility to organic fouling than several other conventional nanofiltration membranes in side-by-side testing.

The new low fouling membranes initially were not included in the side-by-side testing due to capital cost considerations. Hydranautics advised that, as with their low fouling LF series for brackish water treatment, there were additional manufacturing costs associated with producing the low fouling ESNA LF nanofiltration membranes. Hydranautics indicated that the additional manufacturing costs associated with producing the low fouling membrane surface would necessitate a 15 to 20 percent premium over the cost of conventional nanofiltration membranes. For this reason, there was initially a concern that the ESNA LF membrane may not be cost competitive with other nanofiltration membrane elements. However, considering that the membranes from manufacturers B and C did not meet the permeate hardness target and considering that membrane A exhibited troublesome fouling tendencies, it was decided to add the ESNA LF membrane to the side-by-side testing. The results of this testing indicated that the ESNA LF membrane offered the following advantages:

- Much more stable and reliable operation than the Membrane A.

- Lower fouling tendency than the other membranes. The lower fouling tendency should result in reduced cleaning frequency and cleaning costs.

- Lower operating pressure (10 –12 psi lower transmembrane pressure) than the Membrane A. This should result in lower operating costs.

- Better THMFP rejection than Membrane A.

- Ability to meet the permeate hardness requirements.

Most importantly the operational benefits and cost savings associated with the elimination of acid and antiscalant will more than offset the small capital cost premium for these membranes. In addition, through further pilot testing, the properties of the ESNA LF membrane were optimized to meet the treatment City's treatment objectives. The hardness rejection of these membranes was optimized to produce permeate in the City's desired hardness range. This was an important objective for the City since it reduces corrosion control concerns and post-treatment costs and it simplifies operation with respect to blending permeate with lime softened water. These benefits were achieved while maintaining the ability to meet treatment objectives for color and DBPs. In addition to setting a new precedent in terms of nanofiltration capacity, this project introduced several innovations in the approach to pretreatment and in low fouling technology for nanofiltration.

Optimizing New Low Fouling Nanofiltration Membrane Performance
for Deerfield Beach
by

Author: Curtis A. Kiefer – Membrane Treatment Specialist
Camp, Dresser & McKee, Fort Lauderdale, FL

Co-Author: Wayne M. Miller – Chief Operator
City of Deerfield Beach, FL

Co-Author: Jennifer L. Leone – Project Manager
Camp, Dresser & McKee, Fort Lauderdale, FL

Co-Author: Bill Suratt – Consulting Division
Camp, Dresser & McKee, Fort Lauderdale, FL

Introduction

The City of Deerfield Beach, Florida, needed to expand its existing Lime Softening Plant at the West Water Treatment Plant by 10.5 mgd. Membrane Softening was selected as the water treatment process for the West Water Treatment Plant expansion due to the superior water quality which could be achieved and the enhanced ability to meet future projected drinking water regulations. A picture of the Deerfield Beach Membrane Softening Facility under construction is presented in Figure 1. The membrane treatment facility has been designed with five membrane trains, with each train having a capacity of 2.625 mgd. The required 10.5 mgd of capacity to meet projected maximum day demand requirements can be achieved with four trains in operation with the fifth train acting as an installed standby unit.
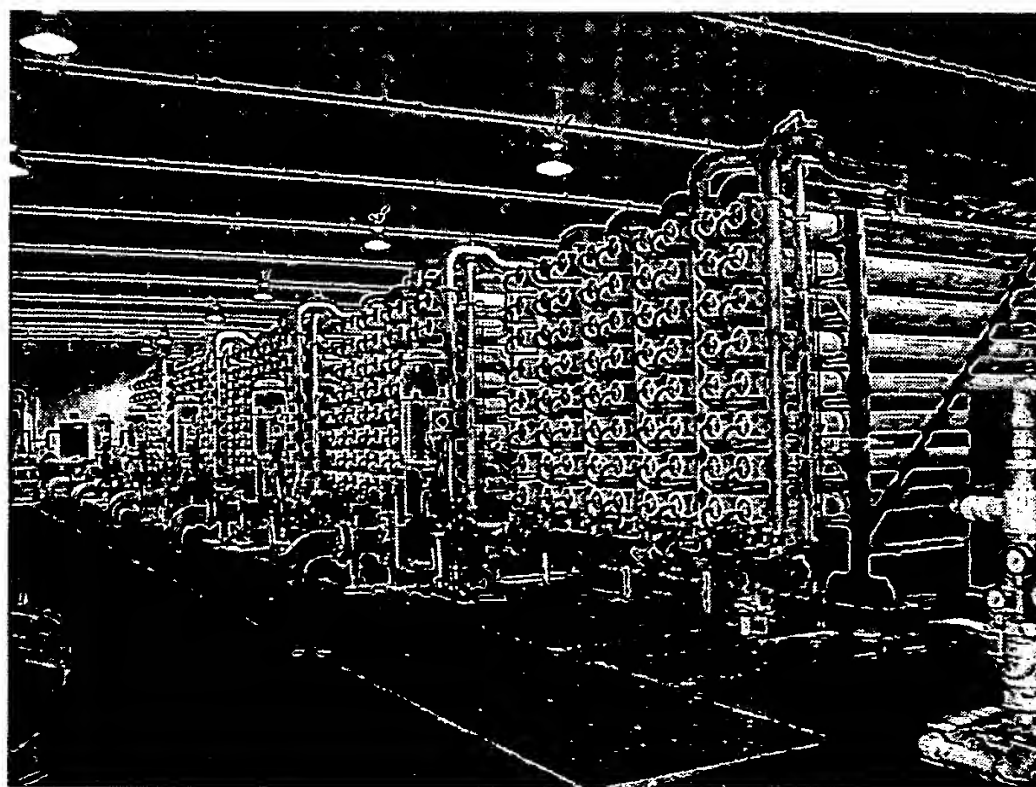

Figure 1

Raw Water Quality and Treatment Objectives

Raw water quality parameters of interest for the City of Deerfield Beach Membrane Softening Plant are summarized in Table 1. The primary water quality parameters which are of concern for this project include hardness, color, DBP's and iron. The average raw water hardness in the City's well water supply is approximately 235 mg/l as calcium carbonate. Several years ago, an American Water Works Association committee set 80 mg/l of hardness as a desirable goal for finished water hardness. The City has been able to produce water with hardness in the range of 70 to 90 mg/l using the lime softening process. However, as discussed below, with the lime softening process the addition of free chlorine to reduce color levels can create challenges in meeting disinfection byproduct regulations. Membrane softening offers the advantage of being able to remove color while at the same time reducing the levels of disinfection byproduct precursors. Membranes are very effective in reducing hardness; however, for corrosion control considerations, the City would prefer to maintain finished water hardness in the range of 70 to 90 mg/l. To achieve this objective, the project specifications require that permeate calcium levels be in the range of 20 to 33 mg/l as calcium ion.

Table 1 Raw Water, Permeate and Finished Water Requirements				
Permeate	Unit	Projected Membrane Raw Water	Required Membrane Permeate	Regulatory Reqs. MCL's
Bicarbonate	mg/l	235	50 to 150	-
Calcium	mg/l	95	10 to 33	-
Total Hardness	mg/l CaCO3	248		-
Chloride	mg/l	33		250
Color	CU	50	<2	15.0
Iron	mg/l	1.5	<0.2	0.30
Sodium	mg/l	25		160
Sulfate	mg/l	34		250
Total Dissolved Solids	mg/l	482	<250	500
Total Trihalomethane Formation Potential	ug/l	400	40	80
Total Haloacetic Acid Formation Potential	ug/l	300	30	60

The City's raw water is generally high in color with values ranging from 30 to 70 color units. Historically the City has reduced color by the addition of up to 20 mg/l of chlorine prior to softening. The addition of high chlorine dosage rates with the relatively high TOC (7.6 mg/l) results in DBP's including total trihalomethanes (TTHM) values that range from 25 to 90 ug/l and total haloacetic acid (THAA) values of 60 to 90 ug/l in the finished water. Since chlorine addition for color removal must be balanced against the

2

need to maintain acceptable DBP levels in the distribution system, color values in the finished water from the lime softening plant typically are in the range of 11 CU. While this color value is below the secondary standard of 15 CU, it is perceptible to a significant number of consumers. For this reason, lowering finished water color levels is an important treatment objective for the City in order to improve customer satisfaction. One of the primary advantages of the membrane treatment process is the ability to remove both color and DBP precursors in a single treatment step. Depending on membrane selection, color should be reduced to 1 to 2 CU in the permeate. In addition, with membrane treatment, THM formation potential (THMFP) and HAA formation potential (HAAFP) should be reduced below the specification requirements of 40 and 30 ug/l respectively.

Another treatment concern for the City is the high levels of iron in the raw water, which historically have averaged 1.6 mg/l with values as high as 2.2 mg/l. The City is rehabilitating four of its wells, which will be dedicated to the membrane softening plant. As shown in Table 1, raw water iron levels for the membrane plant are anticipated to average 1.5 mg/l. Since the rejection of iron should be similar to the rejection of calcium, if higher iron removal is required, the overall permeate hardness will decrease. This means that with high raw water iron levels in the membrane softening process, finished water hardness objectives must be balanced against iron removal objectives. Because there is an MCL for iron but not for hardness, it was considered more important to select the membrane performance to meet the MCL for iron than maintaining a minimum permeate hardness greater than 50 mg/l as $CaCO_3$. In addition, the permeate will be blended with lime softened water in approximately a 60:40 ratio, which will help to keep the finished water hardness in the desirable range. A permeate iron concentration of 0.2 mg/l was established as a permeate treatment objective in order to provide a safety margin on the drinking water MCL of 0.3 mg/l.

Hybrid Membrane Design

At the time that the Deerfield Membrane Treatment plant was originally designed, the Hydranautics ESNA1 membrane was a relatively new nanofiltration membrane available on the market. The manufacturer's specification sheet rated this nanofiltration membrane at approximately 80 percent salt rejection. The corresponding hardness rejection was on the order of 94 percent. Membrane performance projections using the manufacturer's proprietary software program indicated that a membrane system design using the ESNA1 membrane alone would not provide sufficient hardness removal.

To overcome this problem a hybrid membrane design was offered which used higher rejection ESPA1 membranes in the first stage together with the lower rejection ESNA1 membranes in the second stage. The ultra low pressure ESPA1 membranes in the first stage provided the required removal of hardness and other dissolved ions, while installation of the nanofiltration membranes in the second stage resulted in the production of a more balanced finished water and superior flux distribution in the membrane treatment process. This hybrid design took advantage of the lower power consumption

associated with the ESNA1 membrane, while incorporating the higher rejection properties of the ESPA1 membrane to meet the quality objectives.

It was decided that four membrane trains producing 2.625 mgd each should be provided to produce the required permeate capacity of 10.5 mgd. A membrane array of 48:24 was selected for each 2.625 mgd train to achieve an appropriate permeate flux for the system. For the hybrid system design, ESPA1 membranes would be used in the first stage with ESNA1 membranes in the second stage. This configuration provides the most balanced flux distribution for the system and avoids the need for repumping between the first and second stages.

Recent Developments

The nanofiltration market is changing rapidly and the performance of several membrane models has improved since the time that the design was developed for this project. For example, nominal salt rejection of several nanofiltration membranes has increased from 80 percent for the original Deerfield Beach design to over 90 percent at the present time. Hardness rejection of greater than 98 percent has been observed from membranes used for recent pilot testing. Koch has made similar increases in rejection for their 8921 series of nanofiltration membranes. In addition, Dow has made changes to their nanofiltration lineup by replacing their NF70 membrane with a higher rejection NF90 membrane and a lower rejection NF200 membrane. For many membrane softening applications in South Florida, this means that an all nanofiltration design can achieve the required reduction in hardness and iron. An all nanofiltration membrane system design should result in lower operating pressure and lower power consumption as compared to a hybrid design.

Another significant development is the release of a new low fouling nanofiltration membrane. Hydranautics has released a new lower fouling version of the ESNA membrane that has been designated as the ESNA LF (low fouling) membrane. This membrane will use similar low fouling technology, which has already been applied to their LF series of membrane for low pressure RO applications. By lowering the membrane fouling potential, these membranes offer an opportunity to reduce operating costs by lowering power consumption and reducing the cost of chemicals and labor associated with membrane cleaning. It has been observed that naturally occurring organics, predominantly humic acids, have a tendency to foul the membrane surface. As a result of this fouling, the applied pressure to the membranes must be increased to maintain the required permeate flow. This results in an increase in power consumption and the associated operating costs. In addition, more frequent membrane cleaning may be required to control membrane fouling and prevent the fouling from becoming excessive and/or irreversible. By lowering membrane fouling tendency, system reliability will be improved, and system down time and lost production associated with membrane cleaning will be reduced.

Hydranautics has also displayed some flexibility in optimizing the performance of the ESNA LF membrane to meet differing treatment objectives for several softening applications in South Florida.

Low Fouling Membrane Performance Comparison

Mechanisms which affect the rate of membrane fouling by organic molecules include membrane surface charge, charge of the organic molecules, concentration of organics, membrane flux and operating recovery. Pilot plant investigations have demonstrated that different membrane models have different surface charges, which may make that model more susceptible to fouling by organics. Figures 2 and 3 show test results from a side by side test of the ESNA LF membranes versus another nanofiltration membrane (Figure 2) and a hybrid design (Figure 3). Figure 2 illustrates that membrane A experienced a rapid decline in mass transfer coefficient from startup through the first 30 to 40 days of operation. The mass transfer coefficient for membrane A declined by approximately 30 percent while the ESNA LF membrane initially showed a slight increase in mass transfer coefficient and then exhibited very stable operation for the duration of the test. The reference line on the graph indicates the mass transfer coefficient corresponding to the maximum specified membrane operating pressure. If the mass transfer coefficient falls below this value, the maximum specified operating pressure of 80 psi (transmembrane pressure) would be exceeded.
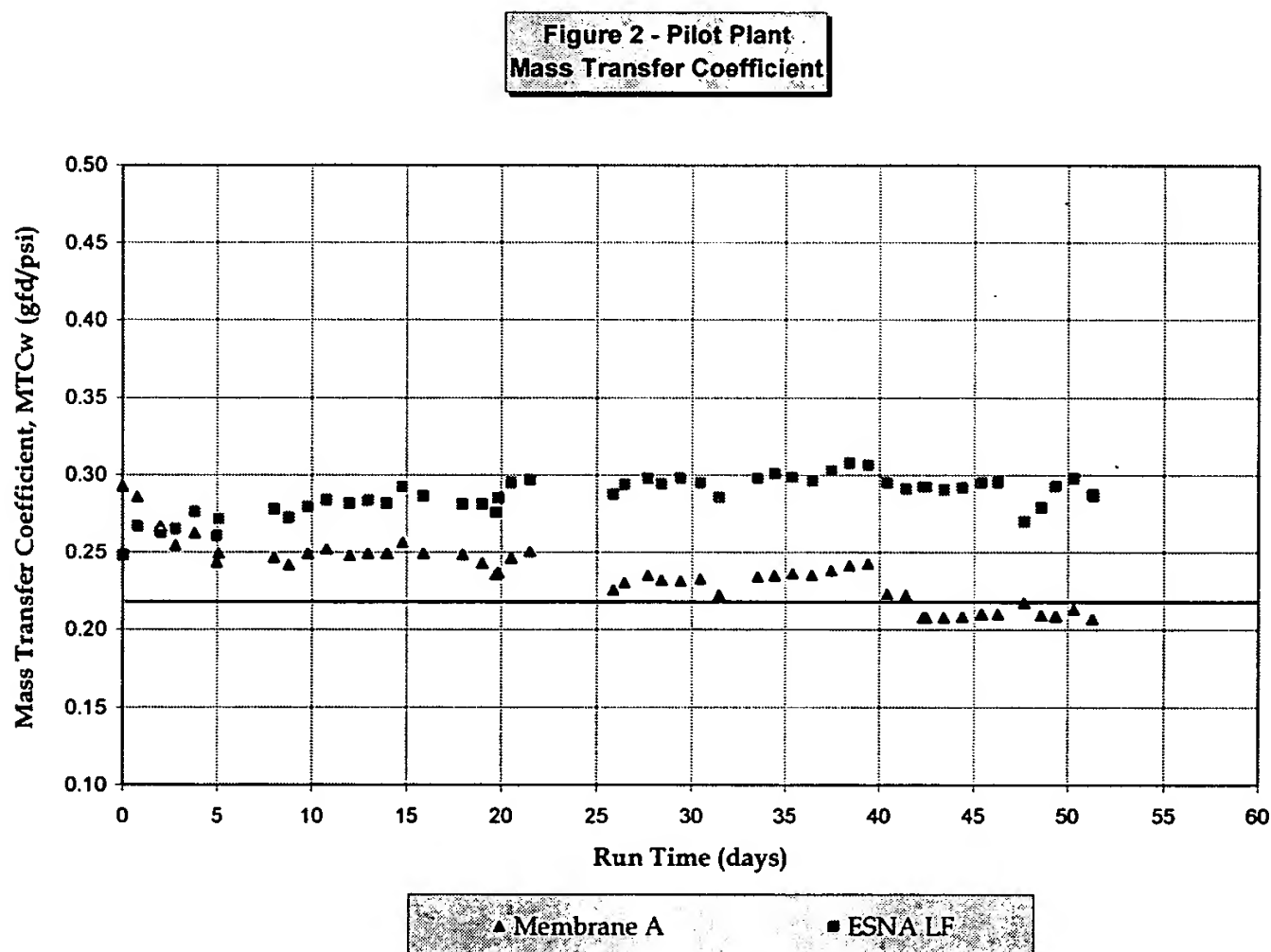

Figure 2 - Pilot Plant Mass Transfer Coefficient

Figure 3 presents a comparison of hardness passage for an ESNA LF system with a Hybrid membrane design similar to the design originally proposed for Deerfield Beach. This hybrid design utilizes an ultralow pressure Membrane B in the first stage and a nanofiltration Membrane A in the second stage. As can be seen from this graph, the hardness passage for membrane A decreased by greater than 50 percent over the first 30 days of testing. The hardness passage of the ESNA LF membrane in similar side by side

5

testing remained relatively constant. As discussed previously, to achieve a maximum iron concentration of 0.2 mg/l in the permeate, permeate hardness will correspondingly be below the minimum desirable range of 50 mg/l as $CaCO_3$. While both the ESNA LF and hybrid designs result in hardness below the desirable range, the permeate hardness for the ESNA LF membrane is closer to the desirable range.

Figures 2 and 3 illustrate the superior operating stability of the ESNA LF membrane. The rapid decline in mass transfer coefficient and hardness passage for membrane A indicates that the performance of this membrane was dramatically affected by fouling. In addition to increased power consumption and cleaning costs associated with this fouling, there is also a concern that cleaning to restore the mass transfer coefficient and lower operating pressure may also significantly increase hardness passage to the initial values. This could result in wide swings in hardness values associated with membrane cleanings.


Figure 3 - Permeate Hardness Low Fouling vs Hybrid
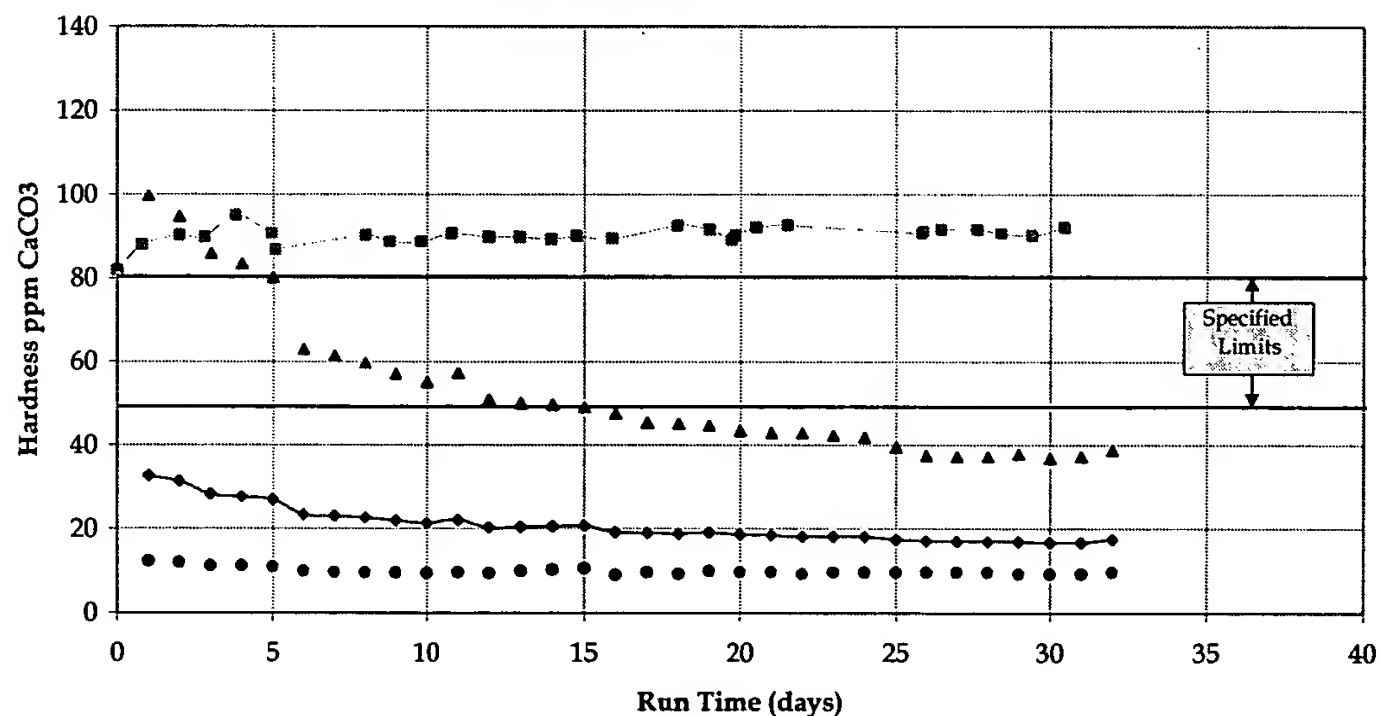
8/31/00

STG 2 - Membrane A   STG 1 - Membrane B   Total   ESNA LF

Optimizing Membrane System Performance

While the testing at other locations demonstrated that the ESNA LF membranes would provide more stable operation than the originally proposed hybrid design, the 92 percent hardness rejection of the ESNA LF membrane reflected in Figure 3 would not be sufficient to reduce the higher iron levels in the Deerfield Beach raw water. As shown in Table 1, the Deerfield Beach raw water iron concentration is 1.5 mg/l while the target permeate concentration was set at 0.2 mg/l. This permeate iron concentration was selected to provide a safety margin on the drinking water MCL of 0.3 mg/l. In order to optimize the ESNA LF membrane design for Deerfield Beach, Hydranautics agreed to

provide their membrane pilot test unit and several sets of membrane elements with varying hardness rejection values.

Figure 4 Deerfield Beach Pilot Permeate Hardness
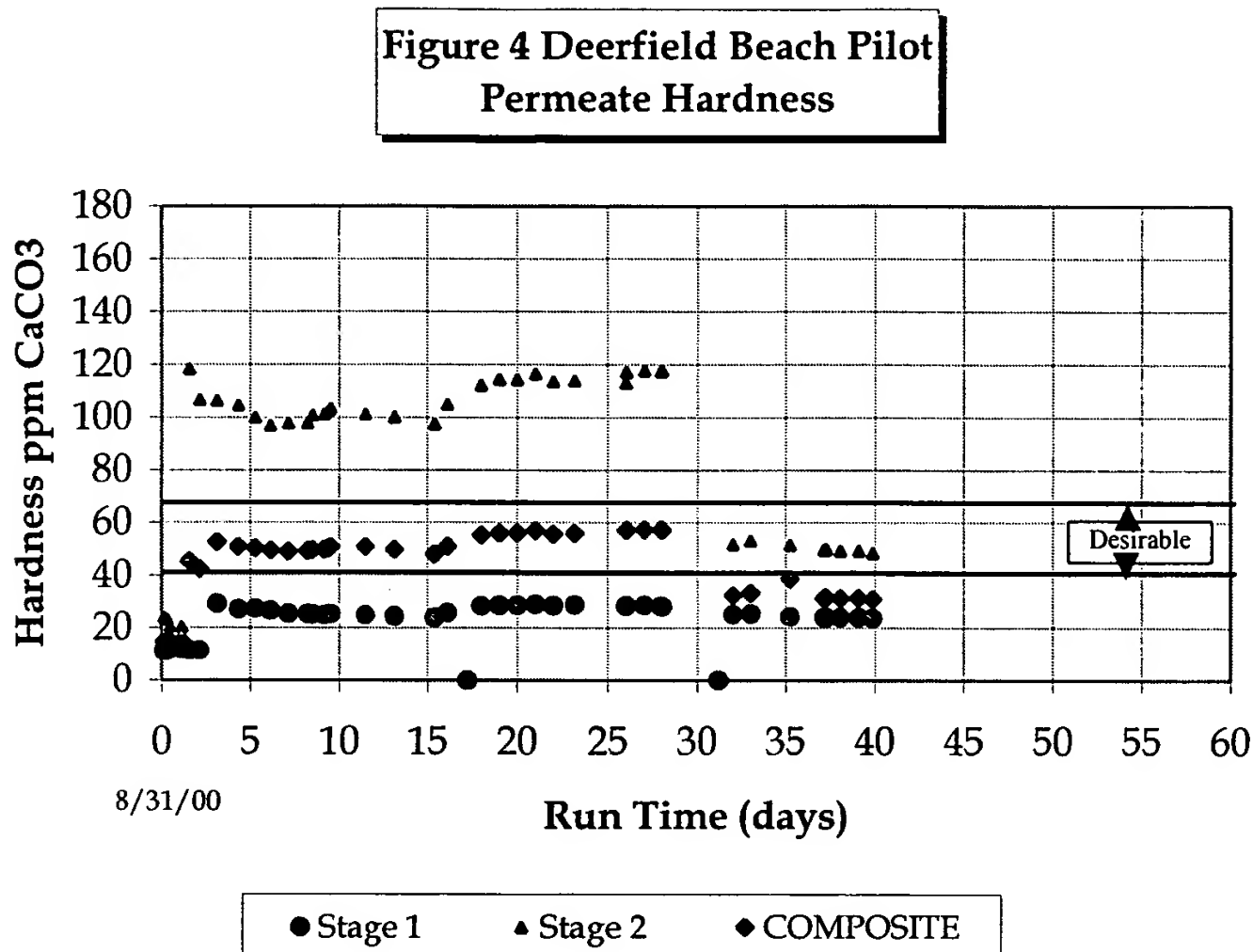
Data from the pilot test runs performed to optimize membrane performance are summarized in Figure 4. Initially the pilot unit was loaded with 98 percent hardness rejection membranes in the first and second stage. As can be seen in Figure 4, permeate hardness values for the first two or three days of operation were less than 20 mg/l. The corresponding permeate iron concentration was approximately 0.03 mg/l. Since these permeate hardness and iron values were much lower than required, it was decided to install 92 percent hardness rejection membranes as the last three elements in each first stage vessel and in the entire second stage of the pilot unit. Operation from day 5 through day 28 reflects the performance of this system configuration. Permeate hardness increased to between 50 to 60 mg/l and permeate iron concentrations increased to between 0.18 to 0.24 mg/l. Overall hardness rejection with this configuration was approximately 94 percent. While these permeate concentrations were generally acceptable, the raw water iron concentration to the pilot unit was only 1.0 mg/l compared to the average raw water concentration of 1.5 mg/l. Because of rehabilitation work on several wells, it was decided to install the pilot unit at one of the well heads that would not be impacted by this work. Unfortunately this well had a lower iron concentration than the average design value. For this reason, higher rejection membrane would be required to meet the permeate quality objectives using raw water with the higher composite iron concentrations.

7

To increase the overall rejection of the system, 95% hardness rejection elements were installed completely in one first stage vessel and the second stage vessel. The other first stage vessel had the same mix of 98 percent and 92 percent hardness rejection membranes as in the previous run. As shown in Figure 4 this modification resulted in a composite permeate hardness of approximately 30 mg/l. This combination also lowered the iron concentrations to 0.11 mg/l at feed pH=6.5. Considering that the iron feed concentrate was approximately 1.0 mg/l for the pilot, permeate iron concentrations could be expected to increase to 0.16 to 0.17 mg/l when operating on composite raw water with iron concentrations of 1.5 mg/l.

Another factor that needed to be considered was feed acidification. As will be discussed in the Economic Analysis section, significant operating cost savings can be achieved by eliminating antiscalant dosing and by reducing acid consumption by raising the acidified feed pH from 6.2 to 6.5; however, if acid dosing could be eliminated completely, even greater cost savings could be realized and safety hazards associated with storage and handling of acid could be avoided. For this reason, it was decided to select the membrane rejection that would result in a permeate iron concentration of 0.2 mg/l with no feed acidification. A higher rejection membrane would be required with no feed acidification because sulfuric acid addition replaces bicarbonate ions with sulfate ions and ion pairs with a cation and sulfate ion are rejected better by the membrane than the same cation paired with a bicarbonate ion. In order to achieve a permeate iron concentration using composite raw water with an iron concentration of 1.5 mg/l with no feed acidification, it was decided to establish the target membrane hardness rejection at 96 percent.

Based on the results of the Deerfield Beach pilot testing, Hydranautics produced a batch of 96 percent hardness rejection membranes. Due to other commitments the pilot unit had to be moved from the Deerfield Beach site before the 96 percent hardness rejection membranes could be tested; however, these membranes were subsequently pilot tested at the Broward County System 1A plant. The raw water quality at the System 1A plant is generally similar to Deerfield Beach. Each plant has hardness in the range of 240 to 250 mg/l, iron in the range of 1.0 to 1.5 mg/l, and color in the range of 50 to 60 CU. As shown in Figure 5, permeate hardness for the System 1A pilot test unit using the 96 percent hardness rejection membranes was in the range of 25 to 35 mg/l. The corresponding permeate iron concentrations are in the range of 0.1 to 0.13 mg/l. The feed iron concentration for this testing was approximately 1.0 mg/l. Extrapolating this data to the Deerfield Beach raw water quality with 1.5 mg/l of iron, the permeate iron concentration should be in the range of 0.15 to 0.20 mg/l. This permeate iron concentration was achieved without feed acidification.

Due to time constraints, it was only possible to operate the Deerfield Beach pilot test without acid and antiscalant addition for approximately 4 days. However, in the subsequent testing at System 1A, all of the pilot testing was conducted without acid or antiscalant addition. As shown in Figure 6, the pilot unit exhibited stable operation without acid and antiscalant addition as evidenced by the stable mass transfer coefficients in the first and second stage.

Figure 5 - Broward County System 1A Hardness
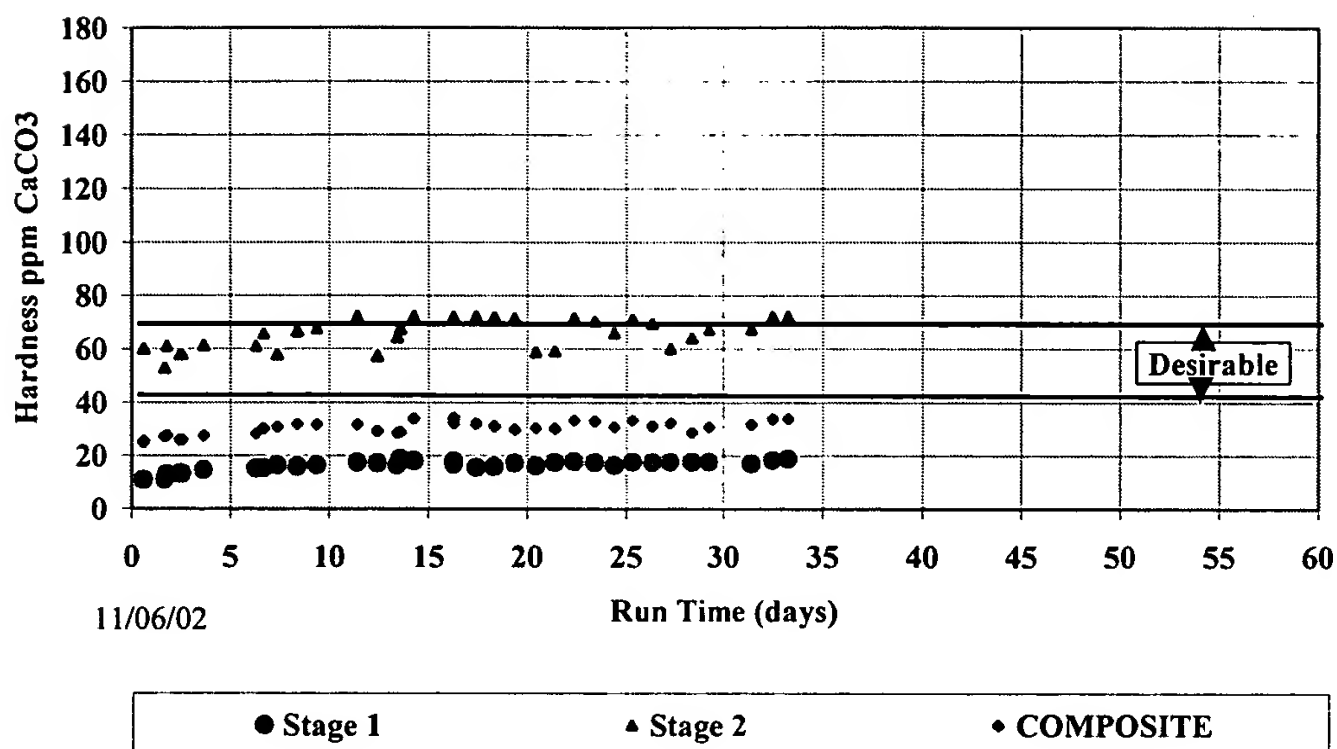
Analytical test results from this testing indicated that the permeate iron concentration was approximately 0.076 mg/l at feed pH=7.2. Unfortunately the corresponding feedwater iron concentration was not analyzed; however, back calculating iron feed concentration from the permeate and concentrate iron concentrations and corresponding flow rates resulted in a feed concentration of 0.8 mg/l. At Deerfield Beach conditions with 1.5 mg/l in the feed, the corresponding permeate iron concentration would be 0.14 mg/l. This value compares with the target permeate iron concentration for Deerfield Beach of 0.2 mg/l with a feed concentration of 1.5 mg/l and a pH=7.0.

Economic Analysis

An economic evaluation was performed to quantify the potential cost savings associated with the use of the low fouling membrane technology compared to the original hybrid membrane design. Potential cost saving areas included lower power consumption, lower pre-treatment chemical consumption, reduced cleaning chemical consumption and cleaning labor. The greatest potential cost savings are associated with reduced pre-treatment chemical consumption.

Reduced Chemical Pretreatment

Based on recently completed pilot and performance testing at Deerfield Beach, Boca Raton and Broward County by CDM, it appears that acid consumption may be reduced and antiscalant dosing eliminated when the low fouling ESNA LF membranes are used. The original hybrid design was based on acidifying the feed to a pH of 6.2 and the

9

addition of 2 to 3 mg/l of antiscalant. The membrane pilot unit at Deerfield Beach operated successfully for over 30 days with feed acidification to a pH of 6.5 without antiscalant addition. Pilot testing at the other locations indicated that both acid and antiscalant pretreatment could be eliminated. Unfortunately, due to other commitments, there was only sufficient time available to conduct a very short run without acid and antiscalant on the Deerfield Beach pilot. For this reason, the cost figures in Table 4 were based on lowering the pH to 6.5 and no antiscalant addition. It is estimated that this reduction in pretreatment chemicals will result in an annual cost savings of approximately $112,400. If acid addition can be completely eliminated, additional cost savings may be realized.

Figure 6- Broward County System 1A
Mass Transfer Coefficient
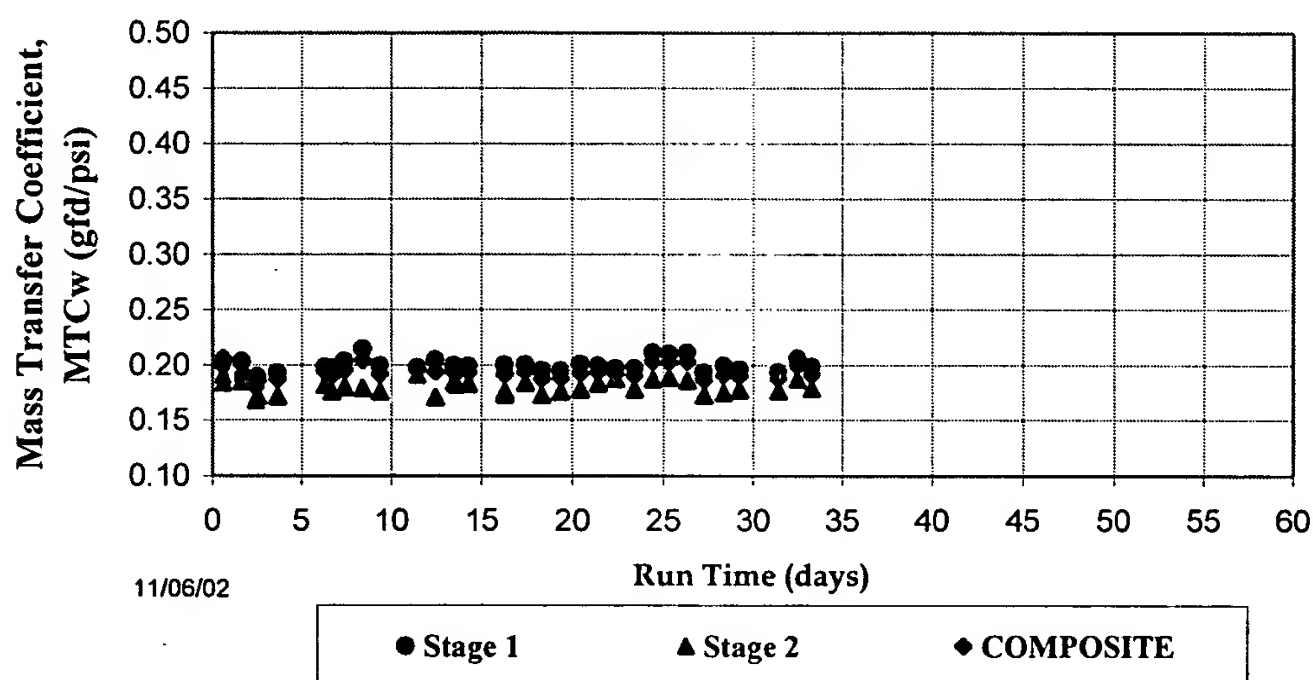
Due to the requirement for reducing permeate iron concentrations to 0.2 mg/l, both systems would reduce permeate hardness to less than 50 mg/l. Both systems would require post-treatment with pH adjustment using caustic soda and the addition of a corrosion inhibitor. For the purposes of this cost estimate it has been assumed that zinc orthophosphate at a dosing rate of approximately 2.5 mg/l of chemical as delivered would be used for corrosion control. Pilot test results indicated that the hybrid system would result in permeate hardness values in the range of 20 to 30 mg/l. When this water is blended at the design blending rate of approximately 60 percent membrane softened with 40 percent lime softened water, the finished water hardness would be less than 40 mg/l. This is much lower than the finished water hardness range of 70 to 90 mg/l preferred by the City for corrosion control. For the low fouling nanofiltration membranes, based on pilot testing and performance projections, it is estimated that with a 60:40 blend of membrane softened with lime softened water, the finished water hardness would be approximately 50 mg/l which is an improvement over the projected hybrid performance. Results of recent pilot testing indicated that the hybrid membrane system may require

10

cleaning approximately every three months while with the low fouling membranes, the cleaning frequency could be reduced to once per year. The savings on cleaning chemicals alone associated with this reduced cleaning frequency would be on the order of $54,000 per year.

Based on pilot testing at another South Florida location, it was originally estimated that power consumption of the high pressure membrane feed pumps could be reduced by almost 30 percent through the use of the low fouling membranes. However, due to the requirement for reducing permeate iron concentration to 0.2 mg/l, higher rejecting ESNA LF membranes will be required at Deerfield Beach compared to the other test location. As a result it is now estimated that only a modest reduction in energy costs will be realized with the low fouling membranes. As shown in Table 4 these cost savings are estimated to be approximately $10,300 per year.

Since the low fouling membranes require an additional treatment step in the manufacturing process to improve the fouling properties of the membrane, these membranes are anticipated to sell at a 15 to 20 percent premium over the cost of conventional nanofiltration membranes. The economic analysis presented in Table 4, includes this additional premium on the initial membrane cost. In addition, the annual membrane replacement cost has been increased proportionately. It is anticipated that this approach may be overly conservative since the cost of new membrane products has shown a downward trend rather than increasing with time, particularly on a cost per thousand gallons basis.

As shown in Table 4, the cost savings associated with the low fouling membranes are anticipated to be on the order of $126,300 per year, even considering the initial additional membrane cost and increased membrane replacement cost. Based on a project life of 20 years, an average daily flow for the membrane plant of 7.6 mgd, a discount rate of 7.5 percent, and an annual inflation rate of 3.5 percent per year, this would translate into a net present value savings of approximately $1.80 million.

In summary, the use of low fouling nanofiltration membranes should reduce pretreatment chemical consumption and should reduce or eliminate the use of sulfuric acid and the associated storage and handling concerns. The associated reduction in the consumption of pretreatment chemicals, cleaning chemicals, and power should result in significant cost savings for the City of Deerfield Beach. Since the membranes for this project are being procured under a separate Membrane Procurement Contract, the City has an opportunity to take advantage of this new technology.

11

TABLE 4 – COST COMPARISON		
Interest Rate :	7.50%	
Plant Life :	20	
Energy Cost $/KwH:	$0.070	
Average Day Flow, mgd:	7.6	
	Membrane Softening Opinion of Cost 2 Stage Design (in $1000's)	
	Low Fouling	**Hybrid**
Plant Capacity (mgd)	10.50	10.50
Recovery Ratio	85.00%	85.00%
Direct Capital Costs		
Sub-Total Direct Costs	11,623.05	11,623.05
Direct Costs ($/gpd)	1.11	1.11
Estimated Indirect Capital Costs		
Interest During Construction, Working Capital, A&E Fee	1,134.90	1,134.90
1. Total Capital Costs + Additional Membrane Cost	$12,946.95	$12,757.95
2. Capital Costs ($/gpd)	$1.23	$1.22
Production Costs (Annual)		
1. Fixed Capital Charges	1269.99	1251.46
2. Power Costs	483.52	493.80
3. Chemicals	243.48	409.50
4. Membrane Replacement	283.50	252.00
5. Replacement Parts & Maintenance	81.44	81.44
6. Direct O & M Labor	190.00	190.00
7. Labor OH & G&A (40%)	76.00	76.00
Total Annual Costs	$2,627.94	$2,754.20
Total Annual Production @ 90% Utilization (Kgal)	2,767,917	2,767,917
Unit Operation Cost (No Capital)($/Kgal)	$0.49	$0.54
Total Unit Production Costs($/Kgal)	$0.95	$1.00
Net Present Value	$31,624.86	$33,427.54
Net Present Value Savings	$1,802.67	-

Note: Capital Costs do not include new generator building or improvements to existing plant

The Deerfield Beach membrane water treatment plant is scheduled to start-up in Spring 2003 with the new Hydranautics ESNA LF membranes installed.

**Advantages of New Low Fouling Nanofiltration Membranes vs.
a Hybrid Membrane Design for Deerfield Beach**
by
**Author: Curtis A. Kiefer – Membrane Treatment Specialist
CDM, Fort Lauderdale, FL**

**Co-Author: Harold E. Jackson – Project Manager
CDM, Fort Lauderdale, FL**

Introduction

The City of Deerfield Beach historically has supplied their customers with lime softened water from their East and West Softening Facilities. The maximum day demand at the time of the water treatment plant expansion feasibility study for this project was 15 million gallons per day (mgd). The permitted capacity of the West Water Treatment Plant (WTP) was 6.0 mgd with the balance of demand being met from the East WTP. At the time of the study, the raw water quality of the wellfield supplying the East WTP had been declining over the past several years due to salt water intrusion. As a result, the City determined that all future water supply should be from the West Wellfield and that the West Plant should be expanded to handle current demand and future projected demand. Projections of future growth indicated that the maximum day demand would increase to 18 mgd by 2010. To meet this demand, it was decided to increase the rating on the existing West Plant from 6.0 to 7.5 mgd. The remaining 10.5 mgd of capacity would be met by a new plant expansion project at the West WTP.

Process options for expansion of the West Water Treatment Plant included lime softening and membrane softening. The lime softening process would be coupled with ozone for color removal and for reducing the formation of disinfection byproducts (DBP's). Membrane Softening was selected as the water treatment process for the West Water Treatment Plant expansion due to the superior water quality which could be achieved and the enhanced ability to meet future projected drinking water regulations. A picture of the Deerfield Beach Membrane Softening Facility under construction is presented in Figure 1. The membrane treatment facility has been designed with five membrane trains, with each train having a capacity of 2.625 mgd. The required 10.5 mgd of capacity to meet projected maximum day demand requirements can be achieved with four trains in operation with the fifth train acting as an installed standby unit.

Raw Water Quality and Treatment Objectives

Raw water quality parameters of interest for the City of Deerfield Beach Membrane Softening Plant are summarized in Table 1. The primary water quality parameters which are of concern for this project include hardness, color, DBP's, and iron. The average raw

water hardness in the City's water supply is approximately 235 milligrams per liter (mg/l) as calcium carbonate. Several years ago, an American Water Works Association (AWWA) committee set 80 mg/l of hardness as a desirable goal for finished water hardness. However, due to higher chemical costs, most softening plants, including the
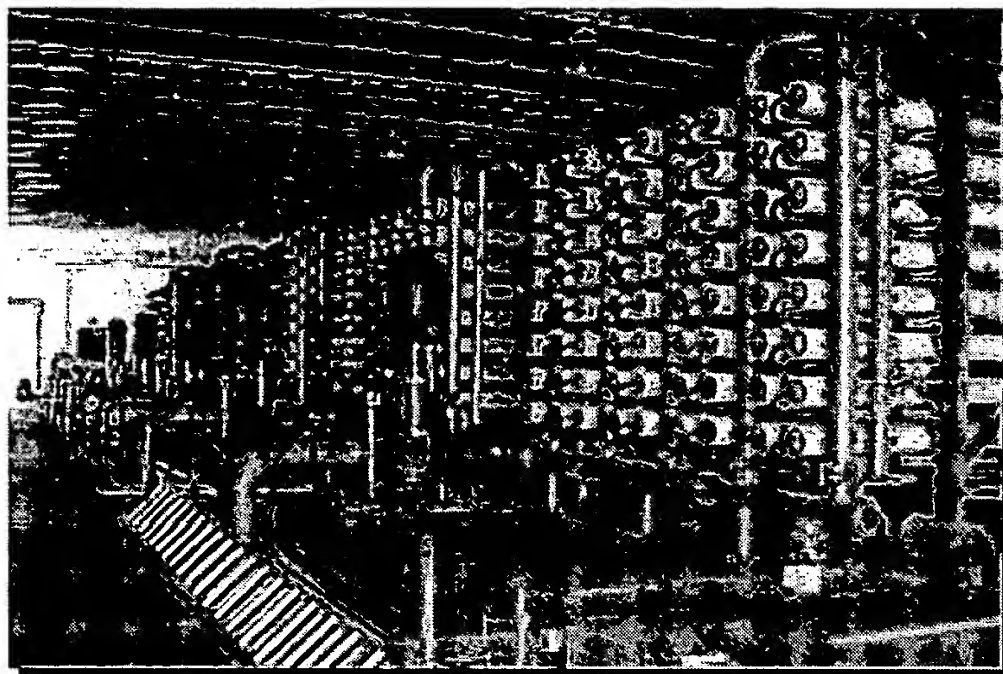

Figure 1

City of Deerfield Beach produces finished water in the range of 80 to 100 mg/l of hardness. The City prefers to maintain finished water hardness in the range of 70 to 90 mg/l. To achieve this objective, the project specifications require that permeate calcium levels must be in the range of 20 to 33 mg/l as calcium ion.

Table 1					
Raw Water, Permeate and Finished Water Requirements					
Permeate	Unit	Projected Membrane Raw Water	Required Membrane Permeate	Projected Finished Water Quality	Regulatory Reqs. MCL's
Bicarbonate	mg/l	235	70 to 150	56.5	-
Calcium	mg/l	95	20 to 33	24.8	-
Total Hardness	mg/l CaCO3	248		65.8	-
Chloride	mg/l	33		33.1	250
Color	CU	50	1	5.4	15.0
Iron	mg/l	1	<0.1	0.16	0.30
Sodium	mg/l	25		18.0	160
Sulfate	mg/l	34		8.7	250
Total Dissolved Solids	mg/l	482	<250	143	500
Total Trihalomethane Formation Potential	ug/l	400	40	25	80
Total Haloacetic Acid Formation Potential	ug/l	300	30	46	60

The City's raw water is generally high in color with values ranging from 30 to 70 color units (CU). Historically the City has reduced color by the addition of up to 20 mg/l of chlorine prior to softening. The addition of high chlorine dosage rates with the relatively high TOC (7.6 mg/l) result in DBP's including total trihalomethanes (TTHM) values that range from 25 to 90 micrograms per liter (ug/l) and total haloacetic acid (THAA) values of 60 to 90 ug/l in the finished water. Since chlorine addition for color removal must be balanced against the need to maintain acceptable DBP levels in the distribution system, color values in the finished water from the lime softening plant typically are in the range of 11 CU. While this color value is below the secondary standard of 15 CU, this level of color is perceptible to a significant number of consumers. For this reason, lowering finished water color levels is an important treatment objective for the City in order to enhance customer satisfaction. One of the primary advantages of the membrane treatment process is the ability to remove both color and DBP precursors in a single treatment step. Depending on membrane selection, color should be reduced to 1 to 2 CU in the permeate. In addition, with membrane treatment, THM formation potential (THMFP) and HAA formation potential (HAAFP) should be reduced well below the specification requirements of 40 and 30 ug/l respectively.

Another treatment concern for the City is the high levels of iron in the raw water, which historically have averaged 1.6 mg/l with values as high as 2.2 mg/l having been reported. The City selected and is in the process of rehabilitating four of its wells, which will be dedicated to the membrane softening plant. As shown in Table 1, with these changes, raw water iron levels for the membrane plant are anticipated to average approximately 1 mg/l. Iron removal will be a significant consideration in membrane selection for this project. Since the rejection of iron should be similar to the rejection of calcium, if a higher rejection membrane is selected to improve iron removal, the overall permeate hardness will decrease. For this reason, with the membrane softening process, finished water hardness objectives must be balanced against iron removal objectives.

Hybrid Membrane Design

At the time that the Deerfield Membrane Treatment plant was designed, the Hydranautics ESNA1 membrane was a relatively new nanofiltration membrane on the market. The manufacturer's specification sheet rated this nanofiltration membrane at approximately 80 percent salt rejection. The corresponding hardness rejection was on the order of 94 percent. Membrane performance projections using the manufacturer's proprietary software program indicated that a membrane system design using the ESNA1 membrane alone would not provide sufficient hardness removal.

To overcome this problem a hybrid membrane design was developed which used higher rejection ESPA1 membranes in the first stage together with the lower rejection ESNA1 membranes in the second stage. The ultra low pressure ESPA1 membranes in the first stage provided the required removal of hardness and other dissolved ions, while installation of the nanofiltration membranes in the second stage resulted in the production of a more balanced finished water and superior flux distribution in the membrane treatment process. This hybrid design took advantage of the lower power consumption

associated with the ESNA1 membrane, while incorporating the higher rejection properties of the ESPA1 membrane to meet the quality objectives. Table 2 presents a performance comparison of the ESPA1 and ESNA1 membranes. As can be seen from this table, the ESNA1 membrane produces 10,000 gallons per day (gpd) of permeate at an applied pressure of 75 pounds per square inch (psi) compared to a permeate production of 12,000 gpd at an applied pressure of 150 psi for the ESPA1. The dramatically lower operating pressure for the ESNA1 membrane would translate into reduced power consumption and lower operating costs.

Table 2 also illustrates the dramatic difference in salt rejection between the two membranes. The ESPA1 membrane has a high salt rejection rate of 99 percent whereas the ESNA1 membrane has a nominal salt rejection of 80 percent. As noted above, at this lower rejection, computer projections indicated that the ESNA1 membrane alone would not be capable of achieving the permeate quality objectives for this project.

Table 2 Membrane Performance Comparison		
Membrane Model	ESPA1	ESNA1
Permeate Flow, gpd	12,000	10,000
Salt Rejection (minimum):	99.0%	80.0%
Hardness Rejection	99.7%	94.0%
Configuration	Spiral Wound	Spiral Wound
Material	Polyamide	Polyamide
Nominal Membrane Area	400 ft^2	400 ft^2
Standard Test Conditions		
NaCl Solution Concentration, mg/l	1500	500
Operating Pressure, psi	150	75
Operating Temperature, °C	25	25
Recovery	15%	15%
PH	6.5 – 7.0	6.5 – 7.0

It was decided that four membrane trains producing 2.625 mgd each should be provided to produce the required permeate capacity of 10.5 mgd. A membrane array of 48:24 was selected for each 2.625 mgd train to achieve an appropriate permeate flux for the system. For the hybrid system design ESPA1 membranes would be used in the first stage with ESNA1 membranes in the second stage. This configuration provides a very balanced flux distribution for the system without repumping between stages.

A comparison of the performance of the membranes in each stage is provided in Table 3. This table illustrates the much higher rejection rate for the ESPA1 elements in the first stage versus the ESNA1 elements in the second stage. The ESPA1 elements were projected to produce permeate with a Total Dissolved Solids (TDS) of only 23 mg/l and a hardness of less than 10 mg/l. In comparison, permeate from the second stage was projected to have a TDS of 488 mg/l and a hardness of 155 mg/l. This would result in a combined permeate with a TDS of 164 mg/l and a hardness of 54 mg/l. This table also

illustrates the well balance flux distribution between the first and second stage with the first stage operating at a flux of 13.6 gallons per ft^2 per day (gfd) compared to a flux of 11.9 gfd in the second stage. Based on this design flux distribution, approximately 70 percent of the total permeate would be produced in the first stage with the second stage producing the remaining 30 percent.

Table 3 Hybrid System Stage by Stage Performance Comparison			
	Stage 1	Stage 2	Total
Capacity, gpm	1,827,816	797,184	2,625,000
No. of Vessels	48	24	72
No. of Elements/Vessel	7	7	-
Area/element, ft^2	400	400	-
Flux, gfd	13.6	11.9	-
TDS, mg/l	22.8	488	164.2
Total Calcium, mg/l	2.8	61	20.4
Total Hardness	9.6	155	53.7
Feed Pressure, psi	75.6	52.7	-
Recovery	59.2%	63.2%	-
Backpressure, psi	0	0	-

Recent Developments

The nanofiltration market is changing rapidly and the performance of several membrane models has changed dramatically since the design was originally developed for this project. For example, Hydranautics has worked to improve the rejection of the ESNA1 membrane and has increased the nominal specification sheet rejection from 80 percent for the original Deerfield Beach design to 90 percent at the present time. Rejection of greater than 94 percent has been observed from membranes used for pilot testing. Koch has made similar increases in rejection for their 8921 series of nanofiltration membranes. In addition, Dow has made changes to their nanofiltration lineup by replacing their NF70 membrane with a higher rejection NF90 membrane and a lower rejection NF200 membrane. For many membrane softening applications in South Florida, an improved rejection rate means a hybrid design is no longer needed to achieve the required reduction in hardness. This will allow municipalities to take full advantage of the lower operating pressure and lower power consumption of nanofiltration membranes, not only in the second stage but throughout the entire membrane array.

In addition, Hydranautics is in the process of releasing a new lower fouling version of the ESNA membrane that may be commercially released as the ESNA LF (low fouling) membrane. This membrane will use unique low fouling technology, in some ways similar to their LFC series of membrane for low pressure reverse osmosis (RO) applications. These lower fouling membranes offer an opportunity to significantly reduce operating
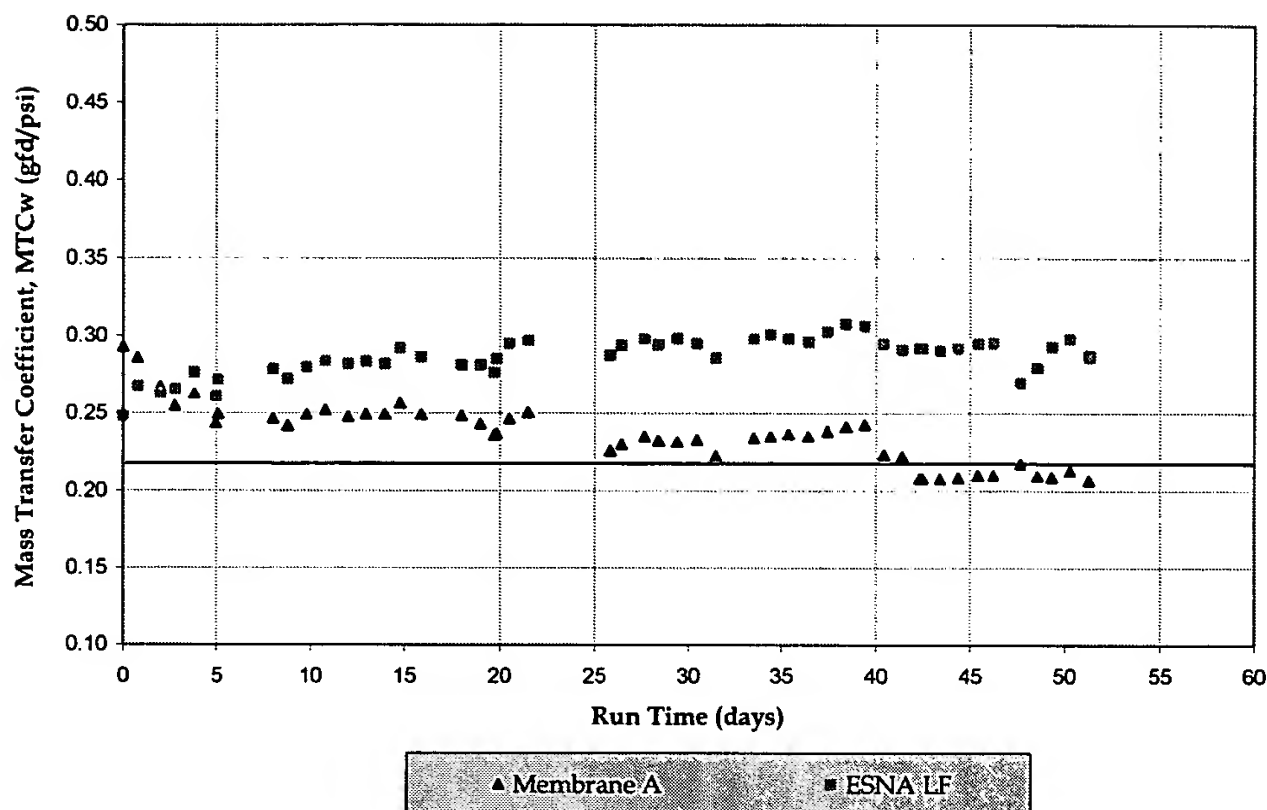
5

costs by lowering power consumption and reducing the cost of chemicals and labor associated with more frequent membrane cleaning. It has been observed that naturally occurring organics have a tendency to foul the membrane surface. As a result of this fouling, the applied pressure to the membranes must be increased to maintain the required permeate flow. Higher applied pressure results in an increase in power consumption and the associated operating costs. In addition, more frequent membrane cleaning may be required to control membrane fouling and prevent the fouling from becoming excessive and/or irreversible. By lowering membrane fouling tendency, system reliability will be improved, and system down time and lost production associated with membrane cleaning will be reduced.

Another benefit of the ESNA LF membrane is that its hardness rejection characteristics have been tailored to provide the optimum hardness rejection for softening applications in South Florida. A number of commercially available nanofiltration membranes were tested side by side in a recent pilot test for another city with similar hardness characteristics as the City of Deerfield Beach. In that testing, the ESNA LF membrane reduced hardness into the range of 85 to 95 mg/l. In this hardness range the permeate is very compatible with the water produced in the lime softening process but with lower color and DBP formation potential. This hardness range is also consistent with the 80 mg/l desirable hardness goal set in the AWWA study mentioned previously. The other commercially available nanofiltration membranes, which were evaluated in the test, tended to produce permeate hardness which was either far below (20 mg/l) or high above (120 mg/l) the optimum range. Permeate with the lower hardness values would either require a higher blending rate of permeate with raw water or lime softened water or would require stabilization with treatment chemicals such as lime or calcium chloride to increase hardness into the desirable range. In a number of applications, additional blending may be problematic due to other limiting factors, such as color and/or disinfection byproduct formation potential. Another alternative would be to accept the lower hardness and use a corrosion inhibitor for corrosion control. The addition of a stabilization chemical and/or a corrosion inhibitor would result in an associated increase in operating cost.

Low Fouling Membrane Performance Comparison

Mechanisms which facilitate membrane fouling by organic molecules include membrane surface charge, concentration of organics, membrane flux, and operating recovery. Previous investigations have demonstrated that different membrane models have different surface charges, which may make that model more susceptible to fouling by organics. Figures 2 and 3 show test results from a side by side test of the ESNA LF membranes versus another nanofiltration membrane (membrane A) and a hybrid design (Figure 3). Figure 2 illustrates that membrane A experienced a rapid decline in mass transfer coefficient from startup through the first 30 to 40 days of operation. The mass transfer coefficient for membrane A declined by approximately 30 percent, while the ESNA LF membrane initially showed a slight increase in mass transfer coefficient and then exhibited very stable operation for the duration of the test. The reference line on the graph indicates the mass transfer coefficient corresponding to the maximum specified membrane
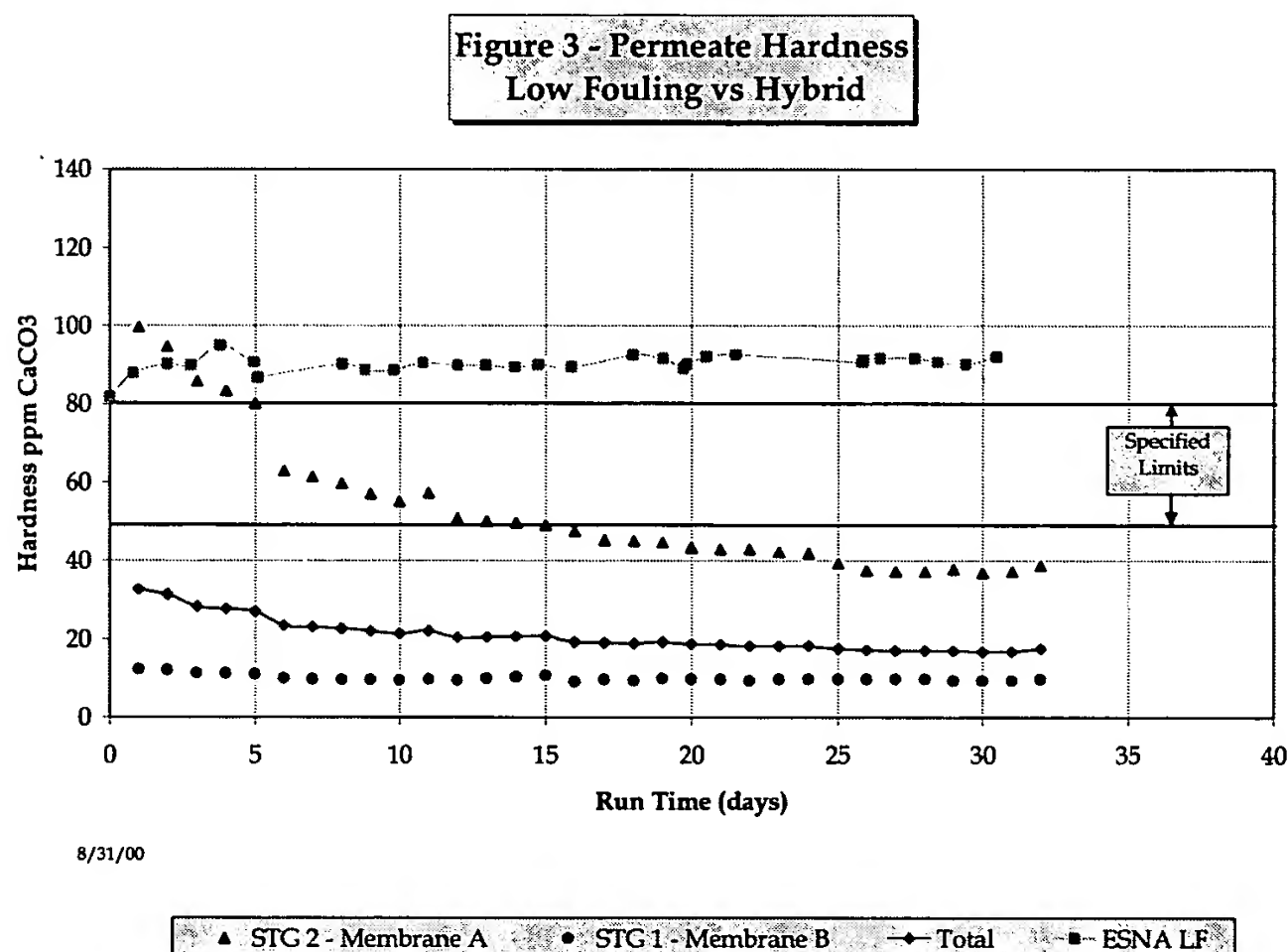
6

Figure 2 - Pilot Plant Mass Transfer Coefficient

operating pressure. If the mass transfer coefficient falls below this value, the maximum specified operating pressure of 80 psi (transmembrane pressure) would be exceeded.

Figure 3 presents a comparison of hardness passage for an ESNA LF system with a Hybrid membrane design using an ultra low pressure Membrane B in the first stage and the nanofiltration Membrane A in the second stage. As can be seen from this graph, the hardness passage for membrane A decreased by greater than 50 percent over the first 30 days of testing. The hardness passage of the ESNA LF membrane in similar side by side testing remained relatively constant. This graph illustrates that the permeate hardness for the Hybrid membrane design would be well below the desirable range of 50 to 80 mg/l which is indicated by the two reference lines on the graph. Permeate produced by the ESNA LF membrane is much closer to the desirable range and is very consistent with the 70 to 90 mg/l of hardness which is typically produced by the lime softening plant. With the low permeate hardness of the Hybrid design, additional post-treatment may be required to stabilize the finished water as discussed previously.

It is important to note that feed acidification has a significant impact on hardness rejection. When sulfuric acid is used for feed acidification, carbonate ions are replaced with sulfate ions. Rejection of the divalent calcium and sulfate ion pair is much greater than for calcium paired with the monovalent bicarbonate ion. In Figure 3, the ESNA LF membrane was tested without acid addition. Computer projections indicate that if the feed is acidified to the original design values for the Deerfield Beach Project, permeate hardness values could be reduced into the 60 to 65 mg/l range. This provides a great deal of latitude in optimizing the permeate and finished water hardness values for the project.

7

Figures 2 and 3 illustrate the superior operating stability of the ESNA LF membrane. The rapid decline in mass transfer coefficient and hardness passage for membrane A indicates that the performance of this membrane was dramatically affected by fouling. In addition to increased power consumption and cleaning costs associated with this fouling, there is also a concern that cleaning to restore the mass transfer coefficient and lower operating pressure may also restore hardness passage to the initial values. This could result in wide swings in hardness values associated with membrane cleanings.

Figure 3 - Permeate Hardness
Low Fouling vs Hybrid

8/31/00

▲ STG 2 - Membrane A   ● STG 1 - Membrane B   ◆ Total   ■ ESNA LF

Economic Analysis

An economic evaluation was performed to quantify the potential cost savings associated with the use of the low fouling membrane technology compared to the original hybrid membrane design. Potential cost saving areas included lower power consumption, lower post-treatment chemical consumption, reduced cleaning chemical consumption, and cleaning labor. Based on recently completed pilot and performance testing for other local South Florida clients, it was estimated that power consumption of the high pressure membrane feed pumps could be reduced by almost 30 percent through the use of the low fouling membranes. Assuming that the membrane treatment plant will be based loaded at a production rate of 10.5 mgd with an annual utilization factor of 90 percent, annual power savings would be on the order of 1.12 million kilowatt hours (kwh)/year. At a power cost of $0.07/kwh, this translates to an annual power cost savings of approximately $78,000 per year.

8

As shown in Figure 3 test results indicated that the hybrid system would result in permeate hardness values in the range of 20 to 30 mg/l. When this water is blended at the design blending rate of approximately 60 percent membrane softened with 40 percent lime softened water, the finished water hardness would be less than 40 mg/l. This is much lower than the finished water hardness range of 70 to 90 mg/l preferred by the City for corrosion control. As mentioned previously, options for increasing hardness to the levels preferred by the City include blending with raw water, blending with lime softened water, increasing the hardness of the lime softened water, chemical addition to increase permeate hardness or addition of stabilization chemicals such as caustic soda in conjunction with a corrosion inhibitor. Since increasing the blending ratio of lime softened to membrane softened water would reduce safety margins on disinfection byproduct formation potential, it was concluded that this was not a desirable solution. Similarly, blending with raw water would increase disinfection byproducts and would increase finished water color. Blending with raw water would practically double finished water color levels, which would negate improvements in color removal associated with the new membrane softening plant. For this reason, raw water blending was not considered to be a viable option. The addition of 19 mg/l of lime to the permeate would result in the lowest cost post-treatment chemical addition scenario. At the base loaded membrane production capacity of 10.5 mgd, this additional chemical consumption would add approximately $29,000 per year in chemical costs.

Results of recent pilot testing indicated that the hybrid membrane system may require cleaning approximately every three months, while with the low fouling membranes, the cleaning frequency could be reduced to once per year. The savings on cleaning chemicals alone associated with this reduced cleaning frequency would be on the order of $54,000 per year.

Since the low fouling membranes require an additional treatment step in the manufacturing process to improve the fouling properties of the membrane, these membranes are anticipated to sell at a 15 to 20 percent premium over the cost of conventional nanofiltration membranes. The economic analysis presented in Table 4 includes this additional premium on the initial membrane cost. In addition, the annual membrane replacement cost has been increased proportionately. It is anticipated that this approach may be overly conservative since the cost of new membrane products has shown a downward trend rather than increasing with time, particularly on a cost per thousand gallons basis.

As shown in Table 4, the cost savings associated with the low fouling membranes are anticipated to be on the order of $128,000 per year, even considering the initial additional membrane cost and increased membrane replacement cost. Based on a project life of 20 years, a discount rate of 7.5 percent, and an annual inflation rate of 3.5 percent per year, the use of low fouling membranes would translate into a net present value savings of approximately $1.83 million.

In summary, the lower fouling membranes appear to offer potential cost savings for the City of Deerfield Beach, while producing a finished water hardness which is more consistent with the City's treatment objectives. Since the membranes for this project are being procured under a separate Membrane Procurement Contract, the City has an opportunity to take advantage of this new technology. A test program is currently underway to confirm that operational benefits will be realized and that the City's treatment objectives will be achieved.

TABLE 4 – COST COMPARISON		
G&A and Profit:	15.00%	
A and E Fee :	6.00%	
Interest Rate :	7.50%	
Plant Life :	20	
Energy Cost $/KwH:	$0.070	
Contractor Eng+OH+Profit :	20.00%	
Utilization Factor:	90.00%	
	Membrane Softening Opinion of Cost 2 Stage Design (in $1000's)	
	Low Fouling	**Hybrid**
Plant Capacity (mgd)	10.50	10.50
Recovery Ratio	85.00%	85.00%
Direct Capital Costs		
Sub-Total Direct Costs	11,582.33	11,582.76
Direct Costs ($/gpd)	1.10	1.10
Estimated Indirect Capital Costs		
Interest During Construction, Working Capital, A&E Fee	1,131.02	1,131.06
1. Total Capital Costs + Additional Membrane Cost	$12,902.35	$12,713.82
2. Capital Costs ($/gpd)	$1.23	$1.21
Production Costs (Annual)		
1. Fixed Charges	1265.62	1,247.13
4. Power Costs	537.33	615.35
5. Chemicals	282.99	366.52
6. Membrane Replacement	283.50	252.00
7. Replacement Parts & Maintenance	81.07	81.07
8. Direct O & M Labor	190.00	190.00
9. Labor OH & G&A (40%)	76.00	76.00
10. G&A and Profit	407.47	424.21
Total Annual Costs	$3,123.97	$3,252.28
Total Annual Production @ 90% Utilization (Kgal)	3,449,250	3,449,250
Unit Operation Cost (No Capital)($/Kgal)	$0.54	$0.58
Total Unit Production Costs ($/Kgal)	$0.91	$0.94
Net Present Value	$38,463.21	$40,293.80
Net Present Value Savings	$1,830.59	-

Note: Capital Costs do not include new generator building or improvements to existing lime softening plant

United States Patent and Trademark Office

Assignments on the Web > **Patent Query**

Patent Assignment Abstract of Title
NOTE: Results display only for issued patents and published applications. For pending or abandoned applications please consult USPTO staff.

Total Assignments: 1

Patent #: 6063278   **Issue Dt:** 05/16/2000   **Application #:** 09067891   **Filing Dt:** 04/28/1998

Inventors: JA-YOUNG KOO, YOUNG SEO YOON

Title: COMPOSITE POLYAMIDE REVERSE OSMOSIS MEMBRANE AND METHOD OF PRODUCING THE SAME

Assignment: 1

Reel/Frame: 009140/0134   **Recorded:** 04/28/1998   **Pages:** 3

Conveyance: ASSIGNMENT OF ASSIGNORS INTEREST (SEE DOCUMENT FOR DETAILS).

Assignors: KOO, JA-YOUNG   **Exec Dt:** 04/27/1998

YOON, YOUNG SEO   **Exec Dt:** 04/27/1998

Assignee: SAEHAN INDUSTRIES INC.
1, JOONGSAN-DONG, KYONGSAN-SI
KYONGSANGBUK-DO, KOREA, REPUBLIC OF

Correspondent: KRIEGSMAN & KRIEGSMAN
EDWARD M. KRIEGSMAN
883 EDGELL ROAD
SUITE 100
FRAMINGHAM, MA 01701

Search Results as of: 04/06/2004 02:34 PM
If you have any comments or questions concerning the data displayed, contact OPR / Assignments at 703-308-9723

| .HOME | INDEX | SEARCH | eBUSINESS | CONTACT US | PRIVACY STATEMENT